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**INNOVATIVE COKE OVEN GAS CLEANING SYSTEM**

**FOR**

**RETROFIT APPLICATIONS**

**FINAL REPORT**

**VOLUME 1: PUBLIC DESIGN REPORT**

**PARTICIPANT**

**Bethlehem Steel Corporation**

**Bethlehem, PA**

**and**

**SUBCONTRACTOR TO BETHLEHEM STEEL CORPORATION**

**Davy/Still-Otto**

**(successor company is Thyssen Still Otto Technical Services)**

**Pittsburgh, PA**

**Prepared for the United States Department of Energy  
Under Cooperative Agreement No. DE-FC22-90PC89658**

**Patents Cleared by Chicago on April 22, 1991.**

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## ABSTRACT

This Public Design Report provides, in a single document, available non-proprietary design information for the "Innovative Coke Oven Gas Cleaning System for Retrofit Applications" Demonstration Project at Bethlehem Steel Corporation's Sparrows Point, Maryland coke oven by-product facilities. This project demonstrates, for the first time in the United States, the feasibility of integrating four commercially available technologies (processes) for cleaning coke oven gas. The four technologies are: Secondary Gas Cooling, Hydrogen Sulfide and Ammonia Removal, Hydrogen Sulfide and Ammonia Recovery, and Ammonia Destruction and Sulfur Recovery. In addition to the design aspects, the history of the project and the role of the U.S. Department of Energy are briefly discussed. Actual plant capital and projected operating costs are also presented.

An overview of the integration (retrofit) of the processes into the existing plant is presented and is followed by detailed non-proprietary descriptions of the four technologies and their overall effect on reducing the emissions of ammonia, sulfur, and other pollutants from coke oven gas. Narrative process descriptions, simplified process flow diagrams, input/output stream data, operating conditions, catalyst and chemical requirements, and utility requirements are given for each unit.

Plant startup provisions, environmental considerations and control, monitoring, and safety considerations are also addressed for each process.

*This Public Design Report describes the processes as of the completion of construction. Any modifications or alterations that occur will be included in the project's Final Report which will be issued following the conclusion of the operations phase.*

Consistent with Bethlehem Steel Corporation's announcement in September of 1991, the production of coke at the Sparrows Point Plant stopped in December of 1991. The coke oven batteries were allowed to "go cold" in January of 1992. The coke oven gas cleaning demonstration facility was prepared for an extended shut-down of at least two years to maintain the facility for a successful start-up if and when coke-making operations are resumed. Bethlehem Steel Corporation is continuing to study a variety of options to resume coke-making at the Sparrows Point Plant. However, it appears unlikely that the coke oven batteries will be operational in the immediate future. Such a decision would be subject to business conditions, availability/demands of funds for other capital projects, and other factors.

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### LIST OF ABBREVIATIONS

BSC	Bethlehem Steel Corporation
CI	Cast Iron
CO	Contracting Officer
COG	Coke Oven Gas
Cond.	Condensate
COTR	Contracting Officer's Technical Representative
CS	Carbon Steel
Dia.	Diameter
DOE	Department of Energy
D/S-O	Davy/Still-Otto
EHSS	Environmental, Health, Safety, and Socioeconomic
EMP	Environmental Monitoring Plan
EPA	Environmental Protection Agency
ETA	Ener-Tech Associates
FL	Flushing Liquor
HP	High Pressure
Ht.	Height
ICCT	Innovative Clean Coal Technology
LP	Low Pressure
MP	Medium Pressure
NEPA	National Environmental Policy Act
PFD	Process Flow Diagram
PON	Program Opportunity Notice
VOC	Volatile Organic Compound
WSAC	Wet Surface Air Cooler

# LIST OF UNITS

BTU	British Thermal Units
BTU/hr	British Thermal Units per hour
BTU/lb	British Thermal Units per pound
BTU/SCF	British Thermal Units per Standard Cubic Foot
Ft	Foot
°F	Degrees Fahrenheit
g	Grams
Gal	Gallons
GPH	Gallons per hour
GPM	Gallons per minute
Gr	Grains
Hp	Horsepower
hr	Hour
KW	Kilowatt
KWH	Kilowatt-hour
l	liter
lb	Pounds
M	Thousand
MM	Million
MMBTU/hr	Million British Thermal Units per hour
MMSCFD	Million Standard Cubic Feet per Day
MMSCFH	Million Standard Cubic Feet per Hour
ppm	Parts per million
PSIG	Pounds per Square Inch Gauge
RPM	Revolutions per Minute
SCF	Standard Cubic Feet
SCFM	Standard Cubic Feet per Minute
SCFH	Standard Cubic Feet per Hour

SCFD                      Standard Cubic Feet per Day

x'-y"                      x feet - y inches

## GLOSSARY OF TERMS

$C_6H_6$	Benzene
$CO_2$	Carbon Dioxide
Deacifier	A trayed column in which acid gases ( $H_2S$ , $HCN$ , and $CO_2$ ) are stripped from the rich liquor by steam and ammonia vapor.
$HCN$	Hydrogen Cyanide
$H_2O$	Water
$H_2S$	Hydrogen Sulfide
$NH_3$	Ammonia
$NH_4HCO_3$	Ammonium Bicarbonate
$(NH_4)_2CO_3$	Ammonium Carbonate
$NH_4CN$	Ammonium Cyanide
$NH_4OH$	Ammonium Hydroxide
$NH_4SH$	Ammonium Hydrosulfide
$(NH_4)_2S$	Ammonium Sulfide

## 1.0 INTRODUCTION

### 1.1 SIGNIFICANCE OF THE PROJECT

The "Innovative Coke Oven Gas Cleaning System for Retrofit Applications" Demonstration Project is the first North American demonstration of an integrated system of commercially proven technologies that can be retrofitted into coke oven gas (COG) handling systems (coal chemical plant) to remove sulfur-bearing compounds and ammonia before the COG is used as a fuel in the steel mill. This innovative COG cleaning technology provides a cost-effective alternative to other COG desulfurization technologies while also reducing the emission rates of several other pollutants.

Traditional COG processing to remove ammonia results in the production of crude ammonium sulfate, a by-product that is very difficult to sell and that requires sulfuric acid to produce. This innovative technology destroys the ammonia, thus eliminating the problems of handling sulfuric acid and disposing of an unmarketable by-product. Other advantages of retrofitting this technology into a typical coal chemical plant are:

- The final cooling system which follows the ammonia saturator will not be required. This eliminates a source of benzene emissions if the existing unit is a water type final cooler.
- The COG pressure drop will be less than through the existing system.
- The ammonia wash desulfurization system has a high turndown capability and also eliminates the need to continuously purchase catalysts or other chemicals for desulfurization.
- The resulting lower COG temperatures will improve light-oil recovery and reduce the naphthalene load on the light-oil system.

Since by-product gas-processing arrangements at the demonstration site are very similar to those of other coal chemical facilities, the retrofitting of this technology should have wide application to the metallurgical coke industry. If this technology reaches full commercialization, sulfur dioxide emissions could be reduced substantially. Based on industry information available at the time of the project development, thirty existing coke plants

in the United States were emitting approximately 300,000 tons of sulfur dioxide each year. This technology could be applicable to twenty-four of these plants and, if economically competitive and installed in all twenty-four plants, sulfur dioxide emissions could be reduced to 40,000 tons per year.

The design capacity of this demonstration facility is 79.5 million SCFD and is based on processing 74 million SCFD of fresh COG and 5.5 million SCFD of recycle tail gas. It is designed to reduce the hydrogen sulfide ( $H_2S$ ) concentration in the COG from 340 Gr/100 SCF to 55 Gr/100 SCF and the ammonia ( $NH_3$ ) concentration from 300 Gr/100 SCF to 4 Gr/100 SCF. This facility is a commercial size unit and would be applicable for retrofit into an above average sized coke-making facility with essentially no scale-up.

Demonstration of this COG cleanup technology is critical to commercialization. When compared to existing COG treatment facilities, the retrofit design is expected to reduce the overall cost of desulfurization, ensure reliable removal of  $H_2S$  from COG, and provide an alternative to overcoming a variety of environmental problems. Demonstration of this technology will show that conventionally accepted equipment and process steps in COG treatment facilities can be eliminated, that retrofit into any existing coke-making facility in the United States is possible without significant downtime, and that COG evolved from coals mined in the United States can be cleaned successfully.

## 1.2 PURPOSE OF THE PRELIMINARY PUBLIC DESIGN REPORT

The purpose of the Public Design Report for the "Innovative Coke Oven Gas Cleaning System for Retrofit Applications" Demonstration Project is to consolidate for public use in one document all available non-proprietary design information that is available at the end of construction. The report also contains background information, an overview of the project, and pertinent cost data.

The scope of the report is limited to non-proprietary information. Therefore, although its content is insufficient to provide a complete tool for designing a retrofit COG cleaning system, it will serve as a reference for the design considerations involved in developing a commercial-scale facility. The report also discusses the advantages of this innovative technology relative to other desulfurization processes.

### 1.3 HISTORY OF THE PROJECT

The Bethlehem Steel Corporation Sparrows Point, Maryland Plant is an integrated steelmaking facility that formerly had a coke-making capacity of approximately 1.4 million tons per year. The coke-making facility had two coal chemicals plants, A and B, which treat the coke oven gas from coke batteries A, 11, and 12 to remove salable constituent chemicals and cleanse the gas for use as fuel. A block flow diagram for Plant A, which treated the gas from Battery A, is shown in Figure 1.3-1 and a block flow diagram for Plant B, which treated the gas from Batteries 11 and 12 is shown in Figure 1.3-2. The batteries are currently on cold idle.

As shown in Figure 1.3-1, the cleaned coke oven gas streams from Plants A and B are combined after light-oil scrubbing. About 60 percent of the combined gas was then desulfurized to meet existing environmental regulations and distributed to consumers within the plant. The remaining 40 percent of the combined gas bypassed desulfurization and was used directly as fuel for underfiring the coke oven batteries.

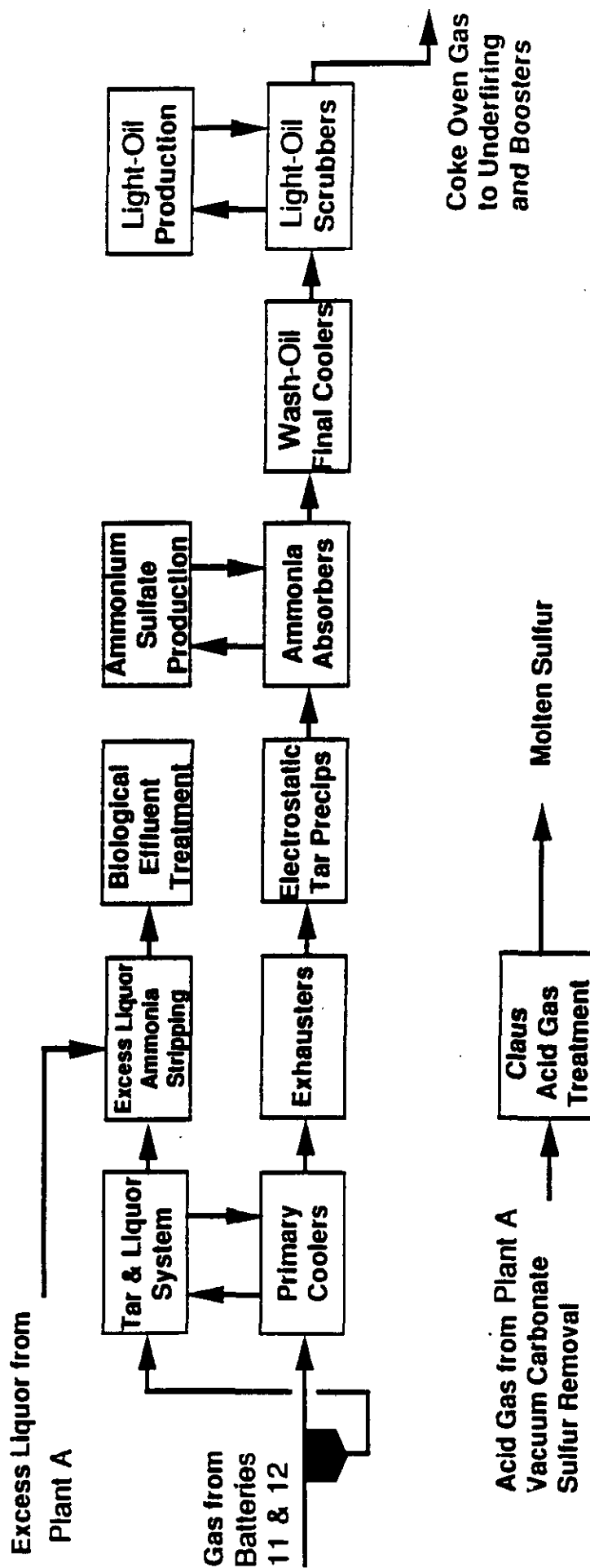
In October of 1987, Bethlehem Steel and the Maryland Department of the Environment signed an Administrative Consent Order that required all coke oven gas to be desulfurized. At that time, inspections of the two coal chemical plants indicated that the existing desulfurization equipment and related facilities for cyanide removal and sulfur recovery had deteriorated to the point where a major rehabilitation/replacement program would be required to maintain acceptable environmental performance for the 60% of the coke oven gas currently being desulfurized. The investment for this renovation program, coupled with the additional investment required for facilities to desulfurize the balance of the gas, plus restoration of other segments in the two coal chemical plants (50 and 30 years old) caused Bethlehem Steel to look for other alternatives that would be capable of cost effectively meeting the environmental regulations.

In order to evaluate the viable alternatives for modifying and improving the Sparrows Point coal chemical plants, Bethlehem Steel commissioned Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) in October of 1987, to study the existing equipment and develop an order of magnitude cost estimate for the rehabilitation of the existing plants. The following criteria were used as the basis for the study: provide operational reliability through to the year 2000; comply with environmental regulations; and investigate process alternatives.

Following a thorough site investigation and discussions with Bethlehem Steel, two main alternatives were identified. These were either a



**Figure 1.3-2**  
**Bethlehem Steel Coal Chemicals Plant "B"**



rehabilitation of the existing coal chemical plants, or major modifications which would involve combining the two coke oven gas streams from Batteries A, 11 and 12 and treating all of the gas in a new desulfurization/ammonia removal system. Installing a combined ammonia removal/destruction and desulfurization system would enhance the feasibility of combining both coal chemical plants. The by-product area would be consolidated and large portions of the old plant equipment would be eliminated. Also, combined desulfurization and ammonia removal using an ammonia wash process would provide the lowest operating cost alternative for removing both the ammonia and hydrogen sulfide.

In order to realistically evaluate the advantages of the proposed modifications, Bethlehem Steel commissioned Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) in February of 1988, to conduct a detailed capital and operating cost comparison of the two alternatives. Plus or minus 10 percent cost estimates were prepared for both alternatives, including the associated benzene emission controls, and a detailed evaluation was made of the respective utility, manpower and maintenance requirements. Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) submitted their report to Bethlehem Steel in April of 1988. The results of this evaluation and a consideration of other tangible and intangible benefits indicated that combining the two coal chemical plants and installing a combined desulfurization/ammonia removal system had significant advantages over the rehabilitation of the existing equipment. In particular, the operating cost savings amounted to approximately 6 million dollars per year while the difference in capital cost was insignificant.

At the time that the study was being developed, the U.S. Department of Energy (DOE) issued a Program Opportunity Notice (PON) for the Second Round of the Innovative Clean Coal Technology Program (ICCT). Under the PON, up to 575 million dollars in federal funding would be provided as cost sharing for projects which demonstrate innovative technology for the reduction of acid rain precursors emitted from plants utilizing coal as a feedstock. A primary emphasis of the PON was to demonstrate innovative technologies that can be retrofitted to existing facilities in the 1990's.

The proposed modifications for Bethlehem Steel's coal chemical plants at Sparrows Point involve innovative technology. A coke oven gas ammonia wash/desulfurization system, whereby all of the ammonia is destroyed at the front of the Claus sulfur recovery facility, has not been installed in the United States. Further, this system has not been combined with a tar and liquor secondary cooling system anywhere in the world. Also, in addition to the reduction in sulfur dioxide emissions, this system would address a

variety of other environmental concerns in a more cost effective manner than systems that have already been employed in the United States.

Since the combined system met many of the important criteria outlined in the PON, Bethlehem Steel and Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) agreed to jointly prepare and submit a proposal to the DOE. A proposal was submitted on May 23, 1988. On September 28, 1988, Bethlehem Steel was notified by DOE that its innovative coke oven gas cleaning system project had been selected as one of 16 projects to enter into negotiations to receive government funding.

Negotiations between Bethlehem Steel and DOE were concluded in the summer of 1989 and Cooperative Agreement Number DE-FC22-90PC89658 was awarded to Bethlehem Steel by DOE on November 14, 1989. Since the time of selection by DOE, Bethlehem Steel, and Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) have been proceeding with the project.

Construction began in February of 1990 and was completed in December of 1991.

Consistent with Bethlehem Steel Corporation's announcement in September of 1991, the production of coke at the Sparrows Point Plant stopped in December of 1991. The coke oven batteries were allowed to "go cold" in January of 1992. The coke oven gas cleaning demonstration facility was prepared for an extended shut-down of at least two years to maintain the facility for a successful start-up if and when coke-making operations are resumed. Bethlehem Steel Corporation is continuing to study a variety of options to resume coke-making at the Sparrows Point Plant. However, it appears unlikely that the coke oven batteries will be operational in the immediate future. Such a decision would be subject to business conditions, availability/demands of funds for other capital projects, and other factors.

## 1.4 THE ROLE OF DOE IN THE PROJECT

### 1.4.1 Innovative Clean Coal Technology Program

In December of 1987, Public Law No. 100-202, as amended by Public Law No. 100-446, provided \$575 million to conduct cost-shared ICCT projects to demonstrate emerging clean coal technologies that can be used to retrofit or repower existing facilities and authorized DOE to conduct the ICCT Program. DOE issued a PON in February 1988 to solicit proposals to demonstrate technologies that were capable of being commercialized in the 1990's, that were more cost effective than current technologies, and that could achieve a significant reduction of sulfur dioxide and/or nitrogen oxides emissions from existing coal-burning facilities, particularly those that contribute to transboundary and interstate pollution.

In response to PON No. DE-PS01-88FE61530, fifty-five proposals were received by DOE in May 1988. All proposals were evaluated by DOE against the six qualification criteria provided in the PON. Proposals that passed the initial qualification review were then evaluated to determine if they met the preliminary evaluation requirements identified in the PON.

For those proposals that remained in the competition, DOE then proceeded to evaluate each offeror's Technical Proposal, Business and Management Proposal, and Cost Proposal. The Technical Evaluation Criteria were divided into two major categories. The first, "Commercialization Factors," addressed the projected commercialization of the proposed technology. The criteria in this section provided for consideration of (1) the potential of the technology to reduce total national emissions of sulfur dioxide and/or nitrogen oxides and to reduce transboundary and interstate air pollution with minimal adverse environmental, health, safety, and socioeconomic (EHSS) impacts; and (2) the potential of the proposed technology to improve the cost effectiveness of controlling emissions of sulfur dioxide and nitrogen oxides when compared to commercially available technology options.

The second major category, "Demonstration Project Factors," recognized that the proposed demonstration project represents the critical step between "pre-demonstration" scale of operation and commercial readiness, and dealt with the proposed project itself. Criteria in this category provided for consideration of the following: the technical readiness for scale-up; the adequacy and appropriateness of the demonstration project; the EHSS and other site-related aspects; the reasonableness and adequacy of the technical approach; and the quality and completeness of the Statement of Work.

The Business and Management Proposal was evaluated to determine the business and management performance potential of the offeror, and was used as an aid in determining the offeror's understanding of the technical requirements of the PON. The Cost Proposal was reviewed and evaluated to assess the validity of the proposer's approach to completing the project in accordance with the proposed Statement of Work and the requirements of the PON.

An overall strategy for compliance with the National Environmental Policy Act (NEPA) was developed by DOE for the ICCT Program, consistent with the Council on Environmental Quality NEPA regulations and the DOE guidelines for compliance with NEPA. This strategy includes both programmatic and project-specific environmental impact considerations during and after the selection process.

In light of the tight schedule imposed by Public Law 100-202 and the confidentiality requirements of the competitive PON process, DOE established alternative procedures to ensure that environmental factors were fully evaluated and integrated into the decision-making process to satisfy its NEPA responsibilities. Offerors were required to submit both programmatic and project-specific environmental data and analyses as a discrete part of their proposal.

The DOE strategy for NEPA compliance had three major elements. The first involved preparation of a comparative programmatic environmental impact analysis, based on information provided by the offerors and supplemented by DOE, as necessary. This environmental analysis ensured that relevant environmental consequences of the ICCT Program and reasonable programmatic alternatives were evaluated in the selection process. The second element involved preparation of a preselection project-specific review. The third element provided for preparation by DOE of publicly available site-specific NEPA documents for each project selected for financial assistance under the PON.

No funds from the ICCT Program would be provided for detailed design, construction, operation, and/or dismantlement until the third element of the NEPA process had been successfully completed. In addition, each Cooperative Agreement entered required an Environmental Monitoring Plan (EMP) to ensure that significant technology, project, and site-specific environmental data are collected and disseminated.

After considering the evaluation criteria, the program policy factors, and the NEPA strategy, DOE selected sixteen proposals for award. On September 28, 1988, Bethlehem Steel was notified by DOE that its innovative coke oven gas cleaning system project had been selected as one of 16 projects to enter into negotiations to receive government funding.

Negotiations between Bethlehem Steel and DOE were concluded in the summer of 1989 and Cooperative Agreement Number DE-FC22-90PC89658 was awarded to Bethlehem Steel by DOE on November 14, 1989.

#### 1.4.2 Management Plan

The DOE entered into a Cooperative Agreement with Bethlehem Steel Corporation, to conduct this project. The DOE will monitor the project through the Contracting Officer (CO) and the Contracting Officer's Technical Representative (COTR). Bethlehem Steel will manage the project through a Program Director, who is assisted by a team of technical and managerial personnel from Bethlehem Steel and from the prime subcontractor, Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services).

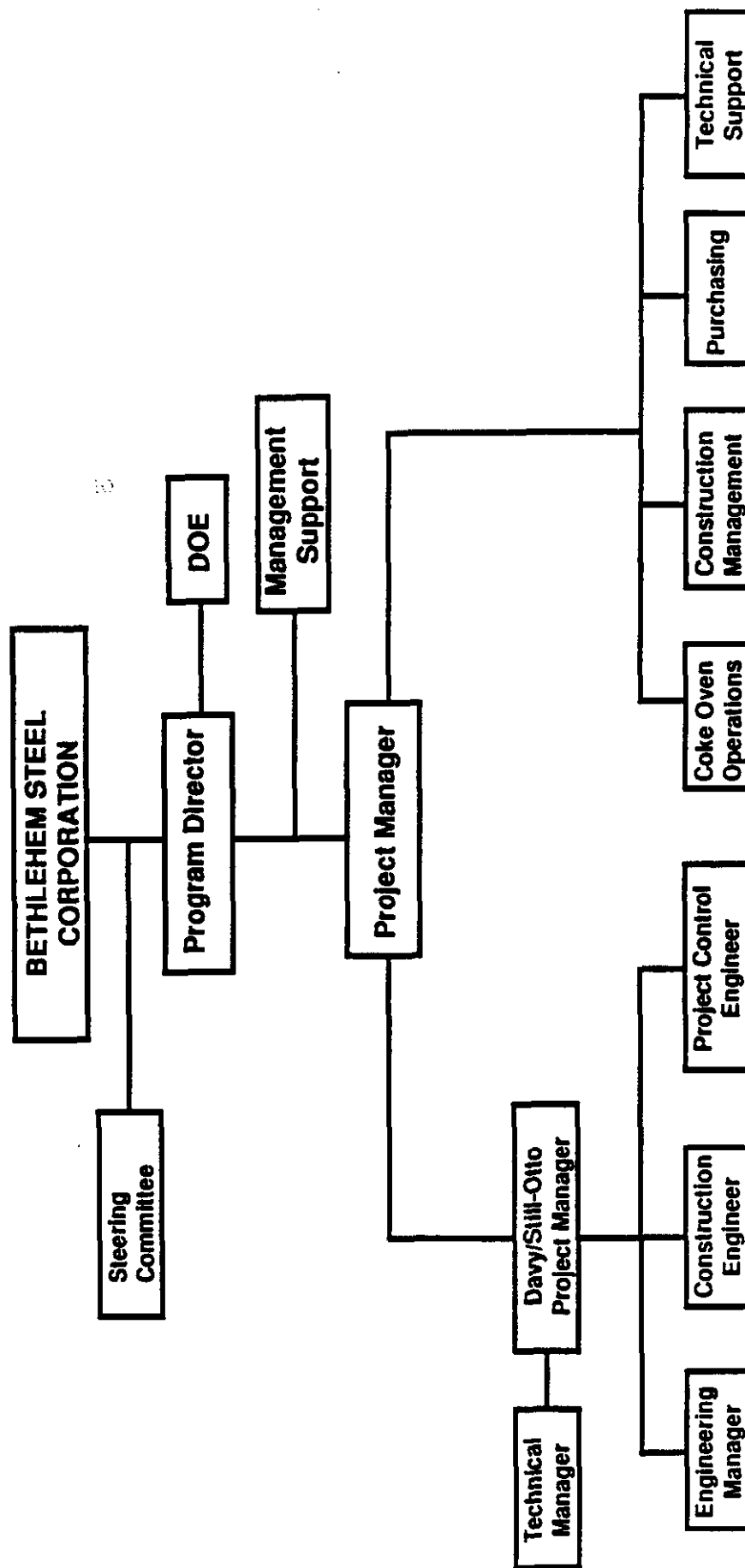
The Bethlehem Steel Corporation organizational chart for this project is shown in Figure 1.4.2-1.

##### 1.4.2.1 DOE

The DOE is responsible for monitoring all aspects of the project and for granting or denying approvals required by the Cooperative Agreement. The DOE CO is the authorized representative of the DOE for all matters related to the Cooperative Agreement.

The DOE CO will appoint a COTR, who is the authorized representative for all technical matters and will have the authority to issue "Technical Advice." The COTR also approves those reports, plans, and technical information required to be delivered by Bethlehem Steel to the DOE under the Cooperative Agreement. The DOE COTR does not have the authority to issue any technical advice that assigns additional work outside the Statement of Work, increases or decreases the total estimated cost or time required for performance of the Cooperative Agreement, changes any of the terms, conditions, or specifications of the Agreement, or interferes with Bethlehem Steel's right to perform the terms and conditions of the Agreement. All Technical Advice will be issued in writing by the DOE COTR.

**Figure 1.4.2-1**  
**Organization Chart for Bethlehem Steel Corporation's**  
**Demonstration Project**



#### 1.4.2.2 Bethlehem Steel Corporation

Bethlehem Steel's Program Director will coordinate the overall project and will be responsible for all communication with the DOE and for interfacing with the DOE COTR. Within Bethlehem Steel Corporation, the Program Director will report to Bethlehem Steel Corporation's Senior Vice President-Operations, who reports directly to the Chairman of the Board and Chief Executive Officer.

A Bethlehem Steel Corporation Steering Committee has been established to provide overall direction for this specific project. The committee, composed of key management personnel, includes the following:

- General Manager, Sparrows Point Plant
- General Manager, Facilities Engineering
- General Manager, Technology
- General Manager, Environmental Affairs
- Manager, Government Programs

The Bethlehem Steel Corporation Steering Committee periodically reviews the progress of the work to verify that key project goals and objectives are being given the priority necessary for success. The committee provides any special guidance as may be necessary to minimize the potential for serious problems and to help in resolving any major problems that may arise.

Bethlehem Steel's Program Director is the primary focal point for this project having responsibility and authority for direction of the project subsequent to the signing of the Cooperative Agreement. He is the principal representative between Bethlehem Steel Corporation and the DOE, and the principal point-of-contact between the DOE, Bethlehem Steel Corporation, and Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) the prime subcontractor.

The Program Director's responsibilities will encompass both technical and fiscal considerations, including the following:

- Overall technical coordination of the program
- Monitoring of program cost
- Monitoring of program planning
- Monitoring of program schedule
- Commitment of resources to optimize performance under the cooperative agreement
- Reporting requirements
- Final review of all deliverables under the cooperative agreement

The Project Manager, designated by Bethlehem Steel Corporation, is responsible for defining design criteria and performance requirements and for continual interactions as necessary to ensure that Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) gets the technical information necessary to support the detailed design of the facility. The Project Manager is also responsible for exercising detailed management controls to see that work is done on schedule and within budget.

For Phase III, Operation, the project team will be kept in place with some modification. During Phase III, Bethlehem Steel's Project Manager will continue to support the Program Director in interactions with Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) in order to promptly resolve any system performance problems and to assist in the evaluation of test results. The Project Manager is also responsible for assuring that system guarantees and warranties are met.

Throughout the course of this project, reports dealing with technical, cost, and environmental aspects of the project will be prepared by Bethlehem Steel and provided to DOE. Bethlehem Steel and Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) will also prepare technical papers describing the proposed technology, its advantages, and the operating results from the demonstration facility to promote the commercialization of the proposed technology for desulfurization, sulfur removal and recovery, and ammonia removal and destruction.

## 2.0 OVERVIEW OF THE PROJECT

### 2.1 INTRODUCTION

The "Innovative Coke Oven Gas Cleaning System for Retrofit Applications" Demonstration Project is the first North American demonstration of an integrated system of commercially proven technologies that can be retrofitted into coke oven gas (COG) handling systems to remove sulfur-bearing compounds and ammonia before the COG is used as fuel in a steel mill. This system was installed at Bethlehem Steel Corporation's Sparrows Point Plant near Baltimore, Maryland. It was designed to process the entire COG stream of 74 million SCFD produced by three coke oven batteries which processed approximately 2 million tons of coal a year to produce 1,400,000 tons of metallurgical coke. The project with an estimated total cost of \$45,239,781 is funded by Bethlehem Steel Corporation (70.2%) and by DOE (29.8%) under Cooperative Agreement No. DE-FC22-90PC89658. Davy/Still-Otto (successor company is Thyssen Still Otto Technical Services) is the prime subcontractor.

The construction of the project started in February of 1990 and was completed in December of 1991.

Consistent with Bethlehem Steel Corporation's announcement in September of 1991, the production of coke at the Sparrows Point Plant stopped in December of 1991. The coke oven batteries were allowed to "go cold" in January of 1992. The coke oven gas cleaning demonstration facility was prepared for an extended shut-down of at least two years to maintain the facility for a successful start-up if and when coke-making operations are resumed. Bethlehem Steel Corporation is continuing to study a variety of options to resume coke-making at the Sparrows Point Plant. However, it appears unlikely that the coke oven batteries will be operational in the immediate future. Such a decision would be subject to business conditions, availability/demands of funds for other capital projects, and other factors.

This project demonstrates, for the first time in the United States, the feasibility of integrating four commercially available technologies (processes) for cleaning coke oven gas. The four technologies are: Secondary Gas Cooling, Hydrogen Sulfide and Ammonia Removal, Hydrogen Sulfide and Ammonia Recovery, and Ammonia Destruction and Sulfur Recovery. The development of each one is discussed separately as follows:

Secondary Gas Cooling - This degree of cooling is commonly used in Europe, where standard operating practices produce lower gas

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The coke-making facility at Bethlehem Steel's Sparrows Point Plant consisted of three batteries of coke ovens, Batteries A, 11, and 12. The COG produced by Battery A was processed in Coal Chemicals Plant A, and the COG produced by Batteries 11 and 12 was processed in Coal Chemicals Plant B. Coal Chemicals Plants A and B will continue to process these two COG streams separately through primary cooling, compression, and tar mist removal.

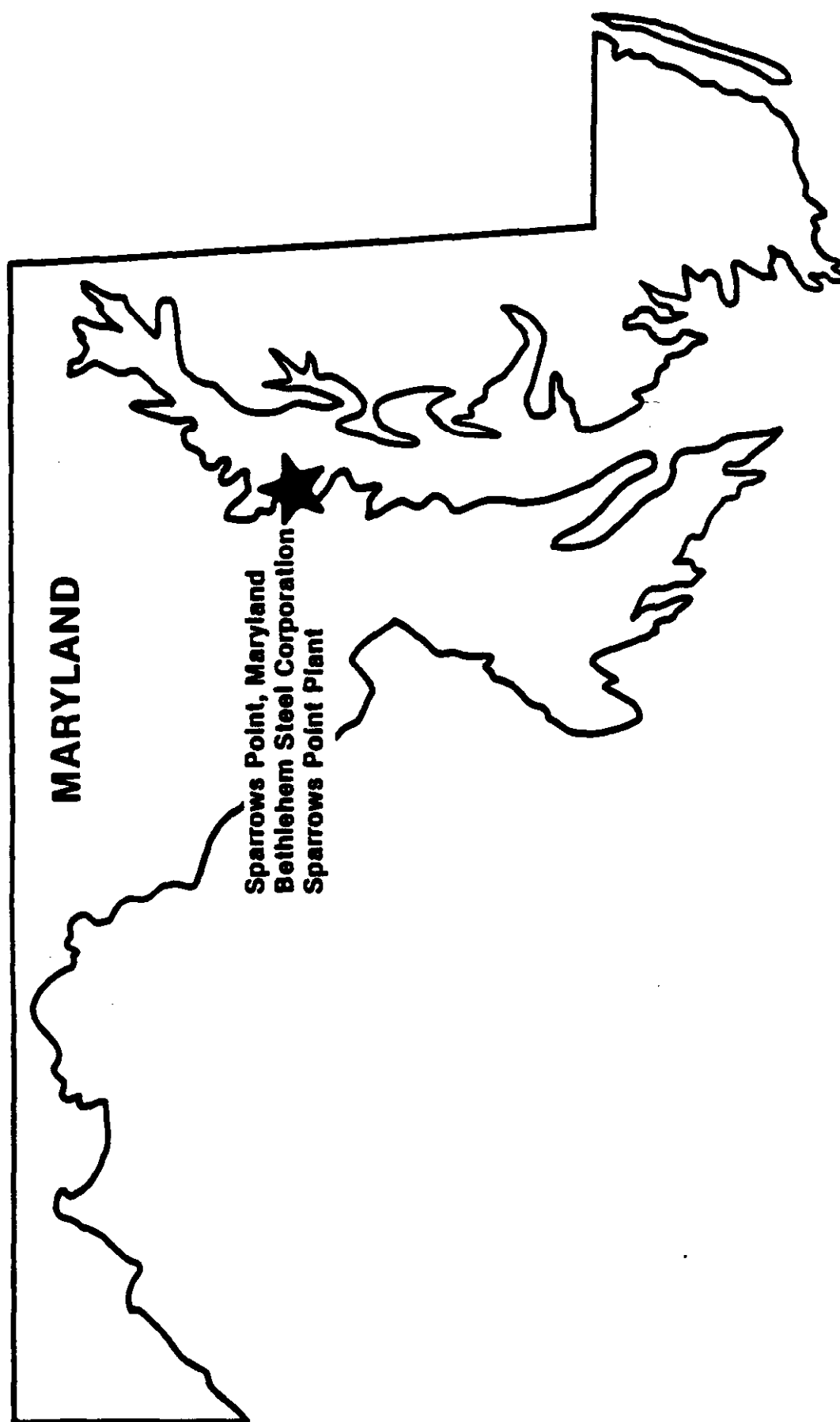
The integrated coke oven gas cleaning process was being retrofitted into Coal Chemical Plant B to treat the combined COG stream from Coal Chemical Plants A and B. The new COG cleaning system will (1) remove and destroy ammonia and hydrogen cyanide and (2) remove hydrogen sulfide and recover a salable sulfur product from the COG. The existing COG cleaning facilities are not as efficient as the demonstration project, produce a difficult to market ammonium sulfate solid by-product, and do not treat the entire COG stream for sulfur removal.

This demonstration project is intended to satisfy the terms of an Administrative Consent Order between Bethlehem Steel Corporation and the State of Maryland, Department of the Environment that requires Bethlehem Steel to desulfurize all COG at its Sparrows Point Plant. This technology is designed to reduce the hydrogen sulfide content of the COG from 340 Gr/100 SCF to 55 Gr/100 SCF and the ammonia content from 300 to 4 Gr/100 SCF. COG desulfurization was mandated by the Maryland Department of the Environment. In addition, emissions of benzene and hydrogen cyanide will be greatly reduced, the ammonium sulfate solid by-product will be eliminated, and a salable sulfur by-product will be produced.

A general description of the integrated process is given in Section 2.4.1 and detailed descriptions of the individual processes as applied in this project are given in Section 3.0.

## 2.2 LOCATION

The demonstration project will be conducted at Bethlehem Steel Corporation's Sparrows Point Plant located in Baltimore County, Maryland (Figure 2.2-1). It is located about 10 miles southeast of downtown Baltimore on a peninsula extending into the Patapsco River. The Patapsco River estuary is also called the Baltimore Harbor. The fully integrated facility to cool the COG, remove and recover hydrogen sulfide and ammonia from the COG, destroy the ammonia, and recover by-product sulfur has been installed at Coal Chemical Plant B which is associated with the coke-making facilities.



**FIGURE 2.2-1 INNOVATIVE COKE OVEN GAS CLEANING SYSTEM FOR  
RETROFIT APPLICATIONS PROJECT SITE**

## 2.3 LAND REQUIREMENTS

The Sparrows Point Plant is an integrated iron and steel complex which occupies approximately 3,000 acres of the Sparrows Point Peninsula. The demonstration project is being retrofitted into the existing Coal Chemical Plant B which occupies 8.6 acres. No additional land, utility or infrastructure is needed. The approximate plot areas required for the new equipment are as follows: Hydrogen Sulfide and Ammonia Scrubbers-65 Ft x 35 Ft; Wet Surface Air Cooling System-60 Ft x 90 Ft; Hydrogen Sulfide and Ammonia Recovery System plus Ammonia Destruction and Claus Plant -60 Ft x 80 Ft.

## 2.4 PLANT FACILITIES

### 2.4.1 General Process Description

#### Existing Facilities

At Coal Chemical Plant A, COG from Battery A is cooled in the two primary gas coolers currently in use. Tar and liquor separated from this gas stream are processed in the three existing flushing liquor decanters. The cooled gas is passed to two of the three existing exhausters, and delivered via the new gas transfer main to the Coal Chemical Plant B area.

At Coal Chemical Plant B, COG from Batteries 11 and 12 is cooled in the two packed primary gas coolers currently in use (Nos. 3 and 4). Recycle hot tail gases from the new Claus sulfur recovery plant, containing residual hydrogen sulfide, will be added to the gas upstream from the primary gas coolers. Tar and liquor separated from the gas stream are processed in the two existing decanters. The cooled gas flow is passed to two of the three existing exhausters and delivered to the tar precipitator inlet header gas main.

The gas stream from Coal Chemical Plant A is combined with the gas stream from Coal Chemical Plant B at the inlet header upstream from the Coal Chemical Plant B tar precipitators. The combined COG stream from Coal Chemical Plants A and B exiting the tar precipitators is directed to the new processing facilities.

The total excess liquor make from Batteries A, 11 and 12 is delivered to the one-million-gallon buffer tank located at Coal Chemical Plant B. New, higher rated pumps are being installed to deliver the liquor from this tank to a new dual vessel sand filter installation. Here, suspended matter, such as tar particles, will be filtered out to reduce potential fouling in subsequent scrubbing and stripping operations.

### New Processing Facilities

The combined COG from the tar precipitators is fed to the lower section of the combined gas cooler/hydrogen sulfide scrubber and flows up through the expanded metal packing counter-current to the descending flow of liquor/tar coolant. The cooled gas then passes upward through the chimney tray into the hydrogen sulfide scrubbing section. The liquor/tar coolant is recirculated from the base of the secondary cooler through dedicated cooling sections of the two combined wet surface air cooler units in which the fluid is cooled to an approach temperature of 5°F to the design wet bulb temperature of 79°F. This provides a cooled gas temperature of 86°F at the inlet to the hydrogen sulfide scrubber.

In the secondary gas cooler additional naphthalene is condensed from the gas stream. The tar present in the circulating liquor coolant dissolves any naphthalene that condenses. A continuous small flow of decanter tar is added to the coolant circuit in sufficient quantity to dissolve the naphthalene. A purge stream of flushing liquor is returned to the Coal Chemical Plant B tar decanters to remove dissolved salts from the cooling circuit and to remove the condensate produced by the secondary cooling operation.

The cooled gas passes upward through the expanded metal packing of the hydrogen sulfide scrubber counter-current to two descending liquid flows: (1) ammonia liquor generated in the ammonia scrubbers that is added at the top of the scrubber, and (2) cooled ammonia-rich deacidified water from the deacidifier unit that is added further down into the scrubber. The scrubber is also provided with two intermediate scrubbing liquor cooling stages that remove the heat of solution generated in the scrubber to maintain the scrubbing efficiency. The liquor is pumped through dedicated cooling sections in the wet surface air coolers and returned to the scrubber.

The COG leaving the top of the hydrogen sulfide scrubber flows to the base of the first of two ammonia scrubbers connected in series. In the ammonia scrubbers, the ammonia present in the COG is absorbed by a counter-current flow of stripped liquor from the free ammonia still. The cleaned COG exits the top of the second ammonia scrubber and is directed to the existing Coal Chemical Plant B light-oil scrubbers.

Excess liquor from storage can also be added to the second ammonia scrubber to reduce the load on the ammonia stills. If used, the excess liquor is first filtered through the new gravel filters and then cooled in a wet surface air cooler unit before entering the scrubber.

The enriched liquor from the hydrogen sulfide scrubber, containing the hydrogen sulfide and ammonia removed from the COG, is pumped from the buffer tank through two sets of plate heat exchangers. In these exchangers, the enriched liquor is preheated by exchange, first with deacidified water, then with ammonia still effluent, up to a temperature of 167°F before entering the deacidifier column. This is a trayed unit in which the acid gases are stripped from the descending liquor by a rising flow of ammonia vapor and steam from the free-ammonia still. The hot, ammonia-rich deacidified water leaving the base of the deacidifier is pumped through the plate heat exchanger and a unit of the wet surface air coolers before being returned to the hydrogen sulfide scrubber for another absorption cycle.

The overhead vapors leaving the top of the deacidifier are a mixture of ammonia, steam, and the acid gases hydrogen sulfide, hydrogen cyanide, and carbon dioxide at a temperature of approximately 187°F. These vapors are cooled to approximately 165°F in a partial condenser to remove some of the water vapor. The condensate is returned to the top section of the deacidifier and the partially cooled acid gas/ammonia stream is passed to the downstream processing units.

A portion of the hot, ammonia-rich deacidified water leaving the deacidifier base is passed to the free-ammonia still where it is stripped of free ammonia and then recycled to the ammonia scrubber for use as the scrubbing liquor.

The free-ammonia still is a trayed column in which the ammonia-rich deacidified liquor is stripped of free ammonia and residual acid gas components. Low-pressure steam, including steam produced in the Claus sulfur recovery unit, is used as the stripping medium. Ammonia-rich vapors are passed from the appropriate intermediate and top sections of the free ammonia-still into the optimum locations in the deacidifier.

The hot stripped liquor leaving the base of the free-ammonia still is divided into two streams. One is a bleed stream that is passed to the fixed-ammonia still to maintain the overall plant water balance. The other is the

main stream that is partially cooled against the deacidifier feed liquor, further cooled in a coil of the wet surface air cooler, and then passed to the ammonia scrubber for use as the absorbing liquor.

The fixed-ammonia still is a trayed column. The feed liquor is intensively stripped of remaining ammonia using the addition of caustic soda solution to release the ammonia which is chemically bound as fixed salts, e.g. chlorides. Low-pressure steam is used as the stripping medium, and the still overhead vapors are passed to an intermediate location on the free-ammonia still. The stripped effluent from the base of the fixed-ammonia still is cooled and pumped to the existing biological effluent treatment plant.

The partially cooled overhead vapors leaving the deacidifier column contain all of the hydrogen sulfide, ammonia, and other acid gases removed from the COG. This acid gas mixture is mixed with air before entering the catalytic reactor which operates at a temperature of 2000°F with a nickel catalyst. In the catalytic reactor, ammonia, hydrogen cyanide, and organic carbon compounds are completely decomposed to produce carbon monoxide, carbon dioxide, nitrogen, hydrogen, and water. The nickel catalyst is insensitive to sulfur compounds at these conditions and, in the presence of water vapor, carbon is not deposited on the catalyst surface. Following the catalytic decomposition, additional air is added to the gas mixture to provide the stoichiometric ratio of oxygen to hydrogen sulfide for the Claus reaction to proceed.

At this point, the first thermal stage of the Claus reaction takes place producing a substantial quantity of sulfur. This sulfur is condensed as the process gases are cooled to approximately 1100°F in a high pressure waste heat boiler. Chemically conditioned and deaerated boiler feedwater is supplied to the boiler which generates 565 PSIG steam. The process gases then pass through a low-pressure waste heat boiler operating at 50 PSIG.

In the low-pressure waste heat boiler, the gases are cooled to approximately 270°F. By controlling a gas bypass around this boiler, low turndown ratios can be achieved for the Claus unit. The process gases leaving the low-pressure waste heat boiler pass in sequence through two catalytic conversion reactors which produce additional sulfur. Sulfur condensers are used to condense the sulfur produced and a process gas heater is used to reheat the gas to the operating temperature of the second catalytic unit. The process gas heater uses steam generated in the high-pressure waste heat boiler. Sulfur condensed from the process gases runs down to a sulfur storage pit.

The tail gases from the Claus unit are returned hot to a point upstream of the Coal Chemical Plant B primary coolers. The control of the Claus unit will allow some hydrogen sulfide to remain in the tail gases and prevent sulfur dioxide from entering the COG. Recycling the tail gas in this manner will still allow overall hydrogen sulfide removal to be maintained at the design value. An emergency thermal oxidizer and stack can be used for periods when the catalytic destruction and sulfur recovery facilities are not in operation.

The COG leaving the new ammonia scrubbers rejoins the existing Coal Chemical Plant B piping at the main gas header at the inlet to the light-oil scrubbers. The gas flows in sequence through one set of three existing scrubbers in series where light oil is absorbed into the circulating wash oil. The light-oil components are stripped from the rich wash oil in the existing Coal Chemical Plant B light-oil recovery system. The primary light oil and secondary light oil continue to be recovered as separate products and are pumped to the new light oil storage tanks. A new product outloading station is also provided. The COG leaving the light-oil scrubbers passes to the existing clean gas distribution system.

The detailed discussion of each of these processes is contained in Section 3.0.

#### 2.4.2 Simplified Block Flow Diagram

Figure 2.4.2-1 represents a simplified Block Flow Diagram of the demonstration plant. In Section 3.0, Input/Output stream data are given for the plant.

#### 2.4.3 Plot Plan/Layout

Figure 2.4.3-1 represents the plot plan for the facility.

#### 2.4.4 Plant Photograph

Figure 2.4.4-1 is an aerial photograph of the plant.

Figure 2.4.2-1

# SIMPLIFIED BLOCK FLOW DIAGRAM OF THE INNOVATIVE COKE OVEN GAS CLEANING SYSTEM

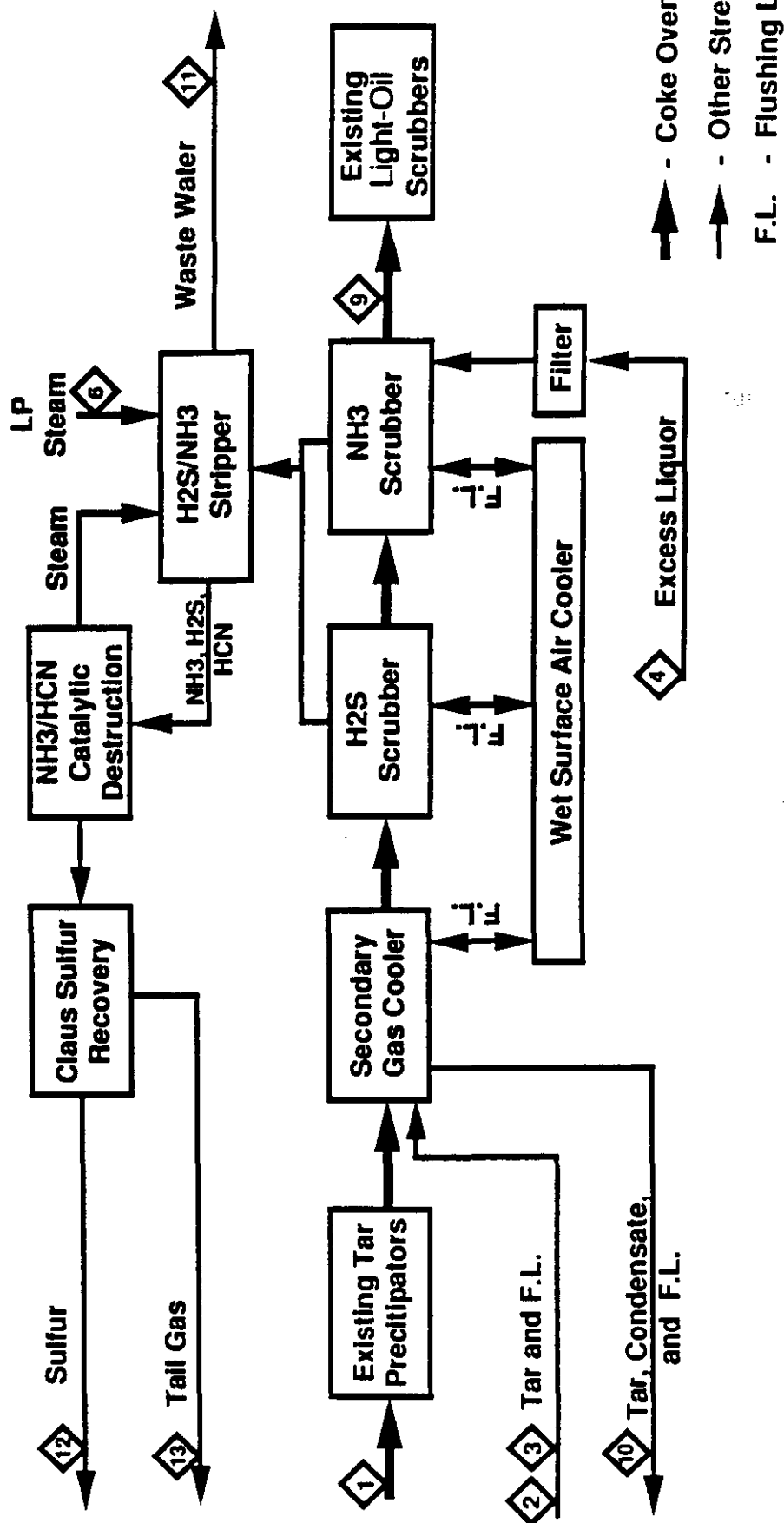






FIGURE 2.4.4-1

DEMONSTRATION  
PLANT SITE

### 2.4.5 Overall Mass Balance and Utility Requirements

#### Mass Balance

The overall material balance corresponding to the input and output streams shown on the Simplified Block Flow Diagram, Figure 2.4.2-1, is summarized as follows (Note: the summary includes streams not shown on Figure 2.4.2-1):

<u>STREAM</u>	<u>COMPONENT</u>	<u>FLOW RATE</u>	
		<u>Volumetric</u>	<u>Mass (lb/hr)</u>
<u>INPUT</u>			
1	Coke Oven Gas	79.5 MMSCFD	112,140
2	Tar	11 GPM	6,050
3	Flushing Liquor	220.5 GPM	110,250
4	Excess Liquor	132 GPM	66,000
5	Boiler Feedwater	19.8 GPM	9,900
6	LP Steam		35,265
7	Air	2172 SCFM	9,970
8	Caustic, 46%	1.95 GPM	1,330
		Total	350,905
<u>OUTPUT</u>			
9	Coke Oven Gas	79.5 MMSCFD	111,700
10	Tar, Cond, and F.L.	247 GPM	123,500
11	Wastewater	220.5 GPM	110,250
12	Sulfur		1,235
13	Tail Gas	2596 SCFM	10,620
		Total	357,305

Note: The mass flow rates (lb/hr) shown above, unless given on the process flow diagrams (PFDs) in Section 3.9, have been estimated from the volumetric flow rates given on the PFDs.

### Utility Requirements

The utility requirements for the new COG cleaning system which include secondary cooling, hydrogen sulfide and ammonia scrubbers, ammonia distillation, deacidifiers, and ammonia destruction and Claus sulfur recovery unit are listed as follows:

<u>Utility</u>	<u>Usage</u>
Electrical Power Absorbed, KW/hr	678
LP Steam, lb/hr	
Consumed	44,065
Generated	8,800
MP Steam Consumed, lb/hr	880
Industrial Water, GPH	27,500
City Water, GPH	1,320
Salt Water, GPH	75,420
Natural Gas, MMBTU/hr	
Normal	0
Startup	1.7

## 2.5 MAJOR PLANT EFFLUENTS

### 2.5.1 Schematic Diagram

A simplified schematic diagram of the major plant effluents is shown in Figure 2.5.2-1.

### 2.5.2 Discussion

During normal operation of the new COG cleaning system, there are no atmospheric emissions from the system itself. However, operation of the new system significantly reduces the sulfur dioxide emissions resulting from combustion of the COG throughout the Sparrows Point Plant. Sulfur dioxide emissions resulting from combustion of the COG processed through the old gas cleaning system were approximately 8900 tons/year. The new system will reduce these sulfur dioxide emissions to about 2600 tons/year or a reduction of approximately 70%. Nitrogen oxides emissions will not be affected since operation of the new system will not alter the heating value or the fuel-bound nitrogen of the product gas.

The new COG cleaning system also minimizes the volatile organic compound (VOC) emissions that are normally emitted from the final cooler. In addition, the installation of the new system results in a significant decrease in fugitive VOC emissions at Sparrows Point because of the shutdown of one of two light-oil recovery units and the replacement of old, leaking equipment with new equipment.

The waste effluents from the new gas cleaning system are wastewater from the ammonia still and blowdown from the boilers and the wet surface air cooler. The ammonia still is designed to discharge 220.5 GPM of wastewater to an existing biological effluent treatment plant. The approximate composition will be 20 ppm hydrogen sulfide, 150 ppm ammonia, 200 ppm carbon dioxide, 10 ppm hydrogen cyanide, and 350 ppm phenols. The blowdown from the boilers and wet surface air cooler will be approximately 120 GPM of wastewater and will be discharged to the sewer system without any treatment.

The new Claus sulfur recovery unit will generate a spent alumina catalyst which will be handled in the same manner as the spent catalyst from the old Claus unit. A spent nickel catalyst will also be generated from the ammonia/cyanide destruction reactor. About five tons of spent nickel catalyst will be generated every 5 to 8 years. This nickel catalyst will be



returned to the vendor for metal recovery or properly managed as a hazardous waste at an off-site treatment or disposal facility.

## 2.6 PRODUCTS/BY-PRODUCTS

### 2.6.1 Description

The main product of the COG cleaning system is a clean fuel gas from which ammonia, sulfur, and other pollutants have been removed. The clean fuel gas exiting the system is first processed to remove light oils and then distributed to consumers within the steel plant. Sulfur is produced as a by-product for sale. Steam is also produced as a by-product but is used within the COG cleaning system.

### 2.6.2 Quantity and Quality

The COG cleaning system is designed to clean 74 MMSCFD of raw coke oven gas from Batteries A, 11, and 12. The composition of the gas entering and leaving the system is as follows:

	% By Volume	
	<u>Raw Gas</u>	<u>Cleaned Gas</u>
Hydrogen	54	52.9
Methane	24	23.8
Nitrogen	9.5	13.3
Carbon Monoxide	6	5.8
Carbon Dioxide	2	2.2
Volatile Organic Compounds	3	2
Ammonia	≤0.9	≤0.02
Hydrogen Cyanide	≤0.1	≤0.05
Hydrogen Sulfide	≤0.5	≤0.09

The COG cleaning system is designed to produce 1235 lb/hr of by-product sulfur from the design flow of raw gas with the above composition. The design flow of steam produced within the Claus unit and used in the ammonia still is 8800 lb/hr.

## 2.7 RELATIONSHIP OF THE PROJECT TO THE SPARROWS POINT PLANT

As discussed in Section 1.3, the Sparrows Point Plant coke-making facility has two coal chemicals plants. Coal Chemical Plant A treats the gas from Battery A and Coal Chemical Plant B treats the gas from Batteries 11 and 12. Several factors have led to Bethlehem Steel's decision to evaluate

viable alternatives for modifying and improving the coal chemical plants. The factors included: (1) The design gas handling capacity of the plants is considerably greater than the present and anticipated gas flowrates and large portions of the plants are old and in need of rehabilitation. (2) Operating costs (operating manpower, maintenance, utility requirements, etc.) are very high for the existing systems. (3) Ammonium sulfate production is no longer an economically viable route for ammonia removal. (4) The installation of benzene emission controls for the present Coal Chemical plant arrangement would be very costly. However, the main factor for developing viable alternatives was that Bethlehem Steel had been mandated by the State of Maryland to desulfurize all of the COG produced, including the gas used for battery heating (See Section 1.3).

The alternative selected by Bethlehem Steel as the best option for achieving compliance with the mandate was to combine the two COG streams from Batteries A, 11 and 12 and treat all of the gas in a new desulfurization/ammonia removal system. Once this option was selected, further evaluations were made to determine the best point to combine the two COG streams. Factors such as the amount of rehabilitation work required at either plant, distances between the two plants, potential condensation problems, etc. led to a decision to treat the gas and condensate in the separate tar and liquor and primary cooling systems of each coal chemical plant and then combine the gas streams on the discharge side of the exhausters.

The relative advantages and disadvantages of both plants for treating the gas after the exhausters were considered. Coal Chemical Plant B offered the following advantages:

- Considerably more space was available at the Coal Chemical Plant B for installing new equipment; space at Coal Chemical Plant A was severely limited by the presence of abandoned equipment.
- Installation of the new equipment in Coal Chemical Plant B would offer a more compact overall layout and minimize the operator requirements and installation costs.
- The light-oil recovery system at Coal Chemical Plant B is more compact, in better shape, and closer to the main gas processing plant.
- The gas mains at Coal Chemical Plant B were in much better condition.

Continuation of light-oil recovery was also evaluated and determined to be beneficial. The Plant B Coal Chemical light-oil system operates well, can handle the total gas flow, and does not require any significant rehabilitation work.

Significant downtime of the COG cleaning system is not expected during construction and start-up of a new system. The new equipment was installed while the existing plant was in operation. Tie-ins to the COG mains were done by hot-tapping. The completed project will allow the Sparrows Point Plant to significantly reduce its sulfur dioxide, ammonia, and volatile organic compound emissions thereby meeting its obligations under the consent order signed with the State of Maryland.

### 3.0 PROJECT PROCESS DATA BY FUNCTIONAL AREA

#### 3.1 PROCESS DESIGN BASIS

The following information has been used as the basis for designing the COG handling system. Most of this information is derived from actual plant operating data. Estimates are provided for the parameters for which data are not available.

Design Gas Rate	Coal Chemical Plant A	-	44.5 MMSCFD
	Coal Chemical Plant B	-	29.4 MMSCFD
Combined Gas Stream		-	74 MMSCFD
Gas Temp. Outlet of Primary Coolers		-	104° F
Gas Temp. Outlet Exhausters		-	122° F
Ammonia in Raw COG		-	300 Gr/100 SCF
H <sub>2</sub> S in Raw COG		-	340 Gr/100 SCF
CO <sub>2</sub> in Raw COG (Estimated)		-	2% Volume
HCN in Raw COG (Estimated)		-	50 Gr/100 SCF
Ammonia in Clean COG		-	4 Gr/100 SCF
Total Sulfur (reported as H <sub>2</sub> S) in Clean COG (55 GR/100 SCF H <sub>2</sub> S and 15 GR/100 SCF Organic Sulfur)		-	70 Gr/100 SCF
Excess Flushing Liquor Rate		-	130 GPM
Analysis - Total NH <sub>3</sub>		-	4250 ppm
Fixed NH <sub>3</sub>		-	1800 ppm
Phenols		-	1000 ppm
HCN		-	50 ppm
H <sub>2</sub> S (Estimated)		-	500 ppm
CO <sub>2</sub>		-	1500 ppm
Typical Plant Design Wet Bulb Temperature		-	79° F

## 3.2 SECONDARY COOLING AND H<sub>2</sub>S SCRUBBING

### 3.2.1 Process Description

#### 3.2.1.1 Secondary Cooling

Reference: Schematic flow diagram in Section 3.2.3.

Since H<sub>2</sub>S absorption is favored by lower temperatures, the gas received from the exhausters is cooled to increase the absorption efficiency. The gas is cooled in the Secondary Cooler to <86°F before it is processed for H<sub>2</sub>S and NH<sub>3</sub> removal. The secondary gas cooling is done in the lower section of the H<sub>2</sub>S Scrubber/Secondary Cooler. In this lower section, the gas enters and flows up through the expanded metal packing countercurrent to the descending flow of liquor/tar coolant. The cooled gas passes upward through the chimney tray into the H<sub>2</sub>S scrubbing section.

The liquor/tar coolant is recirculated from the base of the secondary cooler and through coils of a combined WSAC unit in which the fluid is cooled to an approach of 5°F to the design wet bulb temperature of 79°F. This provides a cooled gas temperature of 86°F at the inlet to the H<sub>2</sub>S Scrubber/Secondary Cooler.

Because the gas is cooled at this stage to a temperature below that achieved in the existing primary gas cooling section, the secondary gas cooler is designed to handle the naphthalene which precipitates from the COG under these conditions. The tar present in the circulating liquor coolant serves this function by dissolving any naphthalene entering the liquor. The coolant circuit includes provisions for a continuous small flow of decanter tar into the system in sufficient quantity to dissolve the naphthalenes. A liquor/tar purge stream is removed from the cooling circuit to prevent accumulation of dissolved salts and to remove gas condensate produced by the secondary cooling operation. A balancing flow of flushing liquor is provided to make up for the purge.

The piping for the secondary cooling liquor circuit is arranged so that, if the H<sub>2</sub>S Scrubber/Secondary Cooler vessel is out of service, the liquor circulation pumps can be used to provide a flow of liquor around the lower section of Ammonia Scrubber 1 to provide standby secondary cooling.

### 3.2.1.2 H<sub>2</sub>S Scrubbing

Reference: Schematic flow diagram in Section 3.2.3.

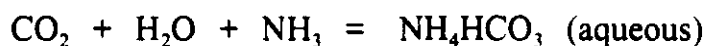
Absorption takes place in the top section of the H<sub>2</sub>S Scrubber/Secondary Cooler where expanded metal packing is used to provide good gas/liquid contact and ease of cleaning. No new chemicals are required since the fluids used to absorb the H<sub>2</sub>S are NH<sub>3</sub>-rich streams obtained from the Ammonia Scrubber and from the Deacidifier. These streams are charged at several column locations. This multi-stage feed of NH<sub>3</sub>-rich liquor streams increases the H<sub>2</sub>S absorption into the liquor by forming NH<sub>4</sub>SH with the absorbed H<sub>2</sub>S. This allows the reaction of H<sub>2</sub>S and H<sub>2</sub>O to proceed.

A portion of the deacidified liquor, containing approximately 25 g NH<sub>3</sub>/liter is cooled in a water-cooled exchanger and fed to a stage near the bottom of the H<sub>2</sub>S Scrubber/Secondary Cooler. Only 65 to 75% of this dissolved NH<sub>3</sub> is available as NH<sub>4</sub>OH for H<sub>2</sub>S scrubbing; the remainder is fixed as (NH<sub>4</sub>)<sub>2</sub>CO<sub>3</sub>, NH<sub>4</sub>HCO<sub>3</sub>, (NH<sub>4</sub>)<sub>2</sub>S, etc.

Strong NH<sub>3</sub> liquor from the Ammonia Scrubbers is fed to the top stage of the H<sub>2</sub>S Absorber and descends the column where H<sub>2</sub>S is absorbed into the NH<sub>3</sub> solution. The degree of absorption is dependent upon the NH<sub>3</sub> and CO<sub>2</sub> concentrations in the solution and the solution/gas temperature. The absorption of H<sub>2</sub>S into the liquid increases as:

- o The solution/gas temperature decreases
- o The ammonia concentration in the liquid increases
- o The CO<sub>2</sub> concentration in the liquid decreases

Some typical reactions are:



Gas from the H<sub>2</sub>S Scrubber/Secondary Cooler flows to Ammonia Scrubber 1 where NH<sub>3</sub> is removed.

The H<sub>2</sub>S Scrubber is provided with two intermediate liquor cooling stages to remove the heat of solution of gases in the scrubber and maintain scrubbing efficiency. This is done by withdrawing liquor from two locations in the H<sub>2</sub>S Scrubber/Secondary Cooler, pumping it through coolers and returning the liquor to the H<sub>2</sub>S Scrubber. Liquor from the H<sub>2</sub>S Scrubber/Secondary Cooler, containing approximately 20 g NH<sub>3</sub>/liter and 5 g H<sub>2</sub>S/liter, is pumped to the Deacidifier for removal of acid gases.

### 3.2.2 Capacity

The unit can operate over a wide range of flows and H<sub>2</sub>S and NH<sub>3</sub> concentrations. The unit will give satisfactory performance at 80 to 100 percent of the design flow rate and with concentrations ranging between 175 and 340 Gr H<sub>2</sub>S/100 SCF and up to 300 Gr NH<sub>3</sub>/100 SCF.

### 3.2.3 Schematic Flow Diagram

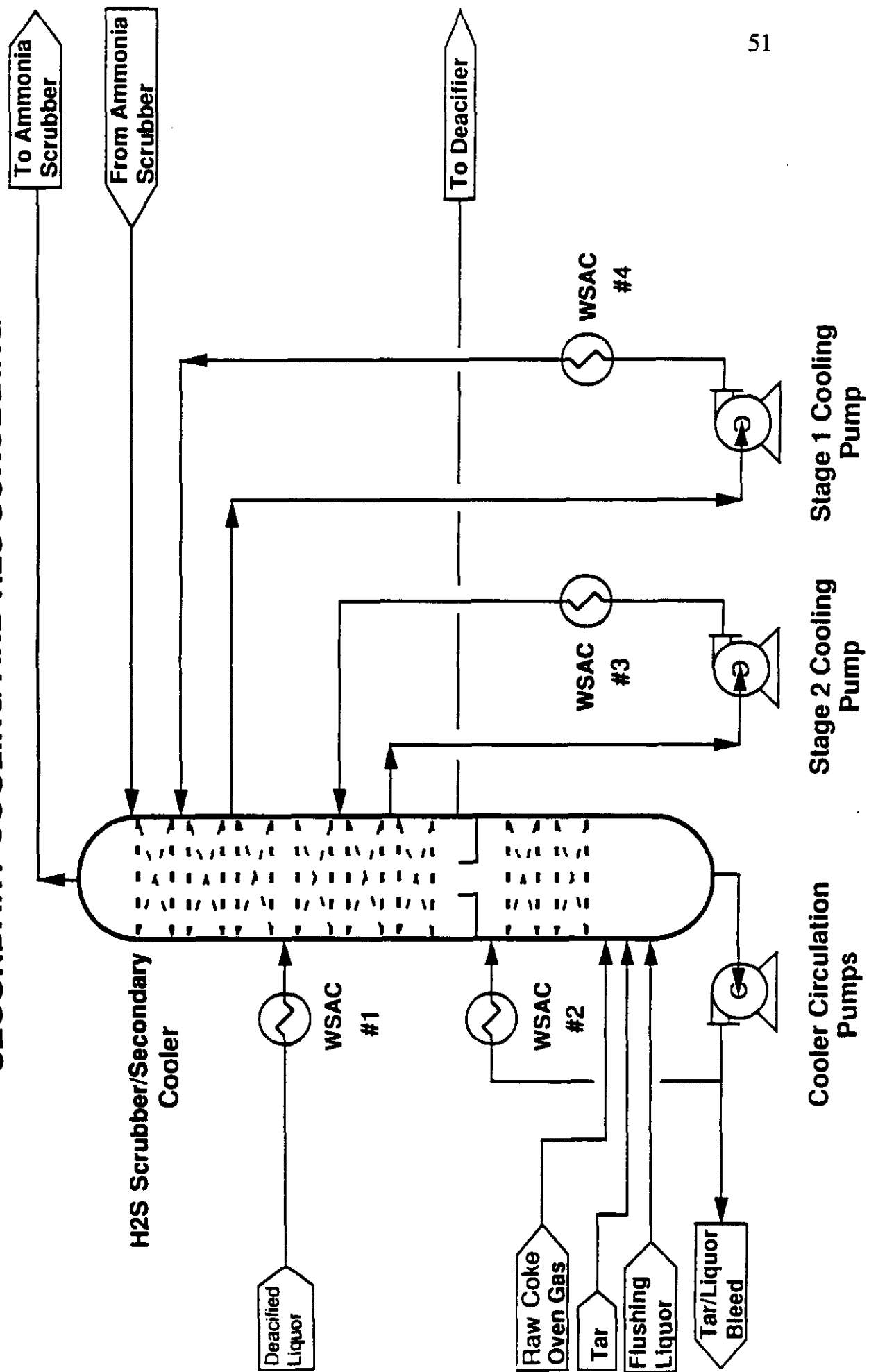
A schematic flow diagram of the Secondary Cooling and H<sub>2</sub>S Scrubbing stage is shown on Figure 3.2.3-1.

### 3.2.4 Major Equipment List and Spares

<u>Equipment No.</u>	<u>Description/Size</u>	<u>Operating</u>	<u>Spare</u>
76.131	#1 Wet Surface Air Cooler 22,500 Ft <sup>2</sup> , SS coil	1	
76.132	#2 Wet Surface Air Cooler 22,500 Ft <sup>2</sup> , SS coil	1	
95.111	H <sub>2</sub> S Scrubber/Cooler 12'-6" Dia. x 112'-9" Ht., CS	1	
95.134	Rich-Liquor Tank 42' Dia. x 30' Ht., CS	1	
95.163.1-2	Cooler Circulating Pumps 1540 GPM, 40 PSIG, 1780 RPM, CI	1	1
95.165.1-2	Rich Liquor Pumps 830 GPM, 65 PSIG, 1770 RPM, CI	1	1
95.166	Stage 2 Cooling Pump 1100 GPM, 35 PSIG, 1770 RPM	1	

Figure 3.2.3-1

# SECONDARY COOLING AND H2S SCRUBBING



<u>Equipment No.</u>	<u>Description/Size</u>	<u>Operating</u>	<u>Spare</u>
95.167	Stage 1 Cooling Pump 1100 GPM, 35 PSIG, 1770 RPM	1	
95.169.1-2	Sump Pumps 100 GPM, 22 PSIG, 1750 RPM	1	1

### 3.2.5 Mass Balance

The mass balance and operating conditions for the Secondary Cooling and H<sub>2</sub>S Scrubbing section of the plant are given on Process Flow Diagram No. 192875 located in Section 3.9.

### 3.2.6 Utility Requirements

Utilities required for Secondary Cooling and H<sub>2</sub>S Scrubbing are listed below. Overall utility requirements for the COG cleaning system, excluding Ammonia Destruction and Claus Sulfur Recovery are shown in Section 3.8.

- o Electricity
- o Cooling water
- o Instrument air
- o Low pressure steam (steam out)
- o Plant air (service)
- o Industrial water (service)
- o Medium pressure steam (service)

### 3.2.7 Catalyst and Chemical Requirements

No purchased chemical are required in this process area. Chemicals for Secondary Cooling and H<sub>2</sub>S Scrubbing are generated within the coke making process, or in the COG clean-up system. The required chemicals are:

<u>Chemical</u>	<u>Use</u>	<u>Source</u>
Flushing Liquor	Direct Contact Gas Cooling	Coking Process
Tar	Dissolve Naphthalene	Coking Process
Ammonia	Increase H <sub>2</sub> S Absorption	COG Cleaning

Purchased water treating chemicals are required for the WSAC. They may include corrosion inhibitors, biocides, pH control, and softening chemicals.

### 3.2.8 Design Considerations

#### 3.2.8.1 Secondary Cooling

**Cooling Fluid** - Direct contact cooling of the COG transfers chemical components from the gas to the liquid phase. Use of a fluid other than one already existing in the process would increase the coolant bleed rate and waste treatment costs. Flushing liquor was selected for this application.

**Naphthalene Deposition** - Quench liquor contains naphthalene. When the quench liquor is cooled in indirect cooling equipment, naphthalene deposits on the cooling surface and reduces heat transfer and increases pressure drop. To avoid this problem, by-product tar is added to the quench liquor stream and keeps the naphthalene in solution.

**Cooling Equipment** - Open loop cooling systems allow contaminants such as  $\text{NH}_3$ ,  $\text{H}_2\text{S}$ , and benzene to escape to the atmosphere. Use of a closed loop cooling system eliminates this problem.

**Tower Packing** - A low pressure drop must be maintained through the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler and the Ammonia Scrubber while providing good contact between the liquor and the gas phases. The packing selected is resistant to plugging and provides for good gas-liquid contact and operates at a low differential pressure.

**Turndown** - Flow rates fluctuate because of the nature of the batch coking process. The scrubbers were designed to operate over a range of flow conditions.

**Equipment Reliability** - To prevent shutdowns or upsets to the process, the equipment must be reliable. Reliability was built into the plant by providing spare pumps and by designing the bottom section of the Ammonia Scrubber to be used as a spare for COG cooling.

**Materials of Construction** - COG contains a variety of chemicals which have the potential to be corrosive under certain environments. Material selection was based on stream composition and operating conditions.

**Emissions** - The system is a closed loop system and normally produces low emissions. The unit was designed for reliable operation to avoid contamination resulting from shutdowns, purging, and equipment cleaning and replacement.

### 3.2.8.2 H<sub>2</sub>S Scrubbing

**Absorption Fluid** - The absorption fluid should be low cost, efficient at low pressure, readily available, not require off-site disposal and be easily regenerated. Use of flushing liquor for absorption meets these requirements. Most commercially available H<sub>2</sub>S absorption systems do not meet these requirements, and some produce a waste stream which must be regenerated or disposed of off-site.

**Turndown** - Flow rates fluctuate because of the nature of the batch coking process. The scrubbers were designed to operate over a range of flow conditions.

**Composition** - The H<sub>2</sub>S, CO<sub>2</sub>, and NH<sub>3</sub> concentrations in both the gas and scrubbing liquor are important design considerations. The H<sub>2</sub>S Scrubber was designed to manipulate these variables to optimize scrubber size. Increased NH<sub>3</sub> concentrations in the liquor increase H<sub>2</sub>S absorption while high levels of CO<sub>2</sub> decrease absorption.

**Temperature** - Operating temperature is an important design consideration. H<sub>2</sub>S absorption is favored by low operating temperature. The final design optimized H<sub>2</sub>S Scrubber size and cooling requirements.

**Tower Packing** - A low pressure drop must be maintained through the H<sub>2</sub>S Scrubber/Secondary Cooler and the Ammonia Scrubber while providing good contact between the liquor and the gas phases. The packing selected is resistant to plugging and provides for good gas-liquid contact and operates at a low differential pressure.

**Contact Time** - Short contact time is desirable to minimize scrubber cost. The design took advantage of the NH<sub>3</sub> concentration and temperature to meet this objective.

**Materials of Construction** - COG contains a variety of chemicals which have the potential to be corrosive under certain environments. Material selection was based on stream composition and operating conditions.

### 3.2.9 Normal Operation

**Reference:** Process Flow Diagram No. 192875 in Section 3.9.

COG is delivered to the Secondary cooler from the Exhauster at a rate of 79.5 MMSCFD, and at a temperature of 129°F. This flow is composed of 74 MMSCFD of raw gas from the coke oven gas batteries, and 5.5

MMSCFD of recycle hot tail gas from the Claus unit. The gas temperature is too high for effective  $\text{H}_2\text{S}$  removal, so it is cooled in the Secondary Gas Cooler. The bottom section of the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler is used for this operation.

COG, Stream 1, enters the bottom of the column and flows upward through the column expanded metal packing where it is contacted and cooled by cold liquid descending through the column. The liquor used to cool the COG is flushing liquor, a by-product from the coking process. Make-up flushing liquor, Stream 15, is fed to the scrubber sump at  $128^\circ\text{F}$ . The COG feed contains naphthalene. To prevent naphthalene condensation in the liquid coolers, a small amount of by-product tar, Stream 14, is injected into the flushing liquor in the bottom of the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler. The Cooler Circulation Pump recycles cooling fluid from the bottom of the scrubber through the WSAC and returns it to the top of the Secondary Cooler where it contacts and cools the gas as the liquid descends through the scrubber's expanded metal packing. The WSAC is a closed system cooling tower that has been installed to eliminate emissions of volatile components such as benzene. Liquid blowdown, Stream 12, is taken at the discharge of the Cooler Circulation Pump and is equal to the liquor fed to the cooler plus COG condensate and tars which are removed in the cooler.

COG is cooled to a temperature of  $86^\circ\text{F}$  passes upward through the Cooler chimney trays and enters the  $\text{H}_2\text{S}$  Scrubber.

Two liquid streams are fed to the  $\text{H}_2\text{S}$  Scrubber. A portion of the deacidifier liquor, Stream 10, containing approximately 25 g  $\text{NH}_3$ /liter is cooled in the Deacidifier Liquor/Rich Liquor and the WSAC exchangers and fed to the middle of the scrubber. Only 65 to 75% of this dissolved  $\text{NH}_3$  is available as  $\text{NH}_4\text{OH}$  for  $\text{H}_2\text{S}$  scrubbing; the remainder is fixed as  $(\text{NH}_4)_2\text{CO}_3$ ,  $\text{NH}_4\text{HCO}_3$ ,  $(\text{NH}_4)_2\text{S}$ , etc.

Strong  $\text{NH}_3$  liquor from the Ammonia Scrubber, Stream 7, is fed to the uppermost stage of the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler. The  $\text{H}_2\text{S}$ -rich liquor containing approximately 20 g  $\text{NH}_3$ /liter and 5 g  $\text{H}_2\text{S}$ /liter is drained to the base of the  $\text{H}_2\text{S}$  Scrubber and collected in the Rich-Liquor Tank. From the Rich-Liquor Tank, the liquor, Stream 13, is pumped by the Rich-Liquor Pump to the Deacidifier.

The  $\text{H}_2\text{S}$  Scrubber is provided with two intermediate liquor cooling stages to remove the heat of solution of gases in the scrubber and maintain scrubbing efficiency. This is done by withdrawing liquor from two

locations in the column, pumping it through dedicated coils in the #1 and #2 WSAC and returning the liquor to the specified column locations.

### 3.2.10 Startup/Upset/Shutdown Conditions

No startup or shutdown problems are anticipated. If scrubber problems are experienced, the COG secondary cooling can be switched to Ammonia Scrubber 1 by setting the proper valves. No additional equipment is required. During this operating mode, a reduction in the  $H_2S$  removal will be experienced which will adversely effect the Ammonia Destruction and Claus sulfur recovery operation.

### 3.2.11 Environmental Considerations

Gaseous emissions from gas cooling - Use of closed loop indirect coolers eliminates the emissions problems associated with direct gas quenching.

Liquid contamination from cleaning equipment - Equipment that has a history of fouling must be opened for cleaning and flushing. Use of non-fouling packing and tar addition to dissolve naphthalene eliminates this problem.

Waste chemical disposal from the absorption process - Many commercial sulfur removal processes generate a hazardous waste stream. Use of an internally generated absorbent eliminates this problem.

Water usage and contamination - Direct gas quenching with water would increase water usage and the amount of water that would have to be treated. Use of closed loop indirect gas coolers eliminates this problem.

By-product production - The absorption step does not generate by-products that must be treated outside the process.

Outside chemical use and disposal - Outside chemicals are required for most commercial  $H_2S$  absorption processes. Since this process uses an internal stream from the coke-making process, no chemicals need to be purchased.

## 3.3 AMMONIA REMOVAL

### 3.3.1 Process Description

Reference: Schematic flow diagram in Section 3.3.3.

The COG leaving the top of the H<sub>2</sub>S Scrubber/Secondary Cooler flows to the base of the first of two Ammonia Scrubbers connected in series. These vessels are designed for a lower gas velocity than the H<sub>2</sub>S Scrubber/Secondary Cooler but have similar expanded mesh packing to facilitate NH<sub>3</sub> removal from the COG to a residual concentration of 4 Gr/100 SCF.

In the Ammonia Scrubbers, the NH<sub>3</sub> present in the COG is absorbed by a countercurrent flow of stripped liquor from the Free-Ammonia Still. Provisions are made to allow the additional use of excess liquor for scrubbing, thus reducing the load on the Ammonia Stills. Provisions are also made for supplementing the flow of stripped liquor for scrubbing with softened potable water. This is required only during abnormal conditions, such as Ammonia Still failure. The softening equipment required for the boiler feedwater duty includes extra capacity for this purpose.

Ammonia Scrubber 1 is equipped to serve as a standby H<sub>2</sub>S Scrubber/Secondary Cooler. The bottom section acts as standby for the secondary cooling duty. The use of a specially designed chimney tray above the bottom section enables enriched liquor to flow via a seal pot to the enriched liquor buffer tank, thus keeping the secondary cooling liquor in the bottom section separate from the enriched liquor. In this way, tar in the secondary cooling liquor is not allowed to contaminate the enriched liquor.

While using Ammonia Scrubber 1 as a standby H<sub>2</sub>S Scrubber/Secondary Cooler, the H<sub>2</sub>S removal is not as efficient. However, NH<sub>3</sub> removal is maintained at the design efficiency.

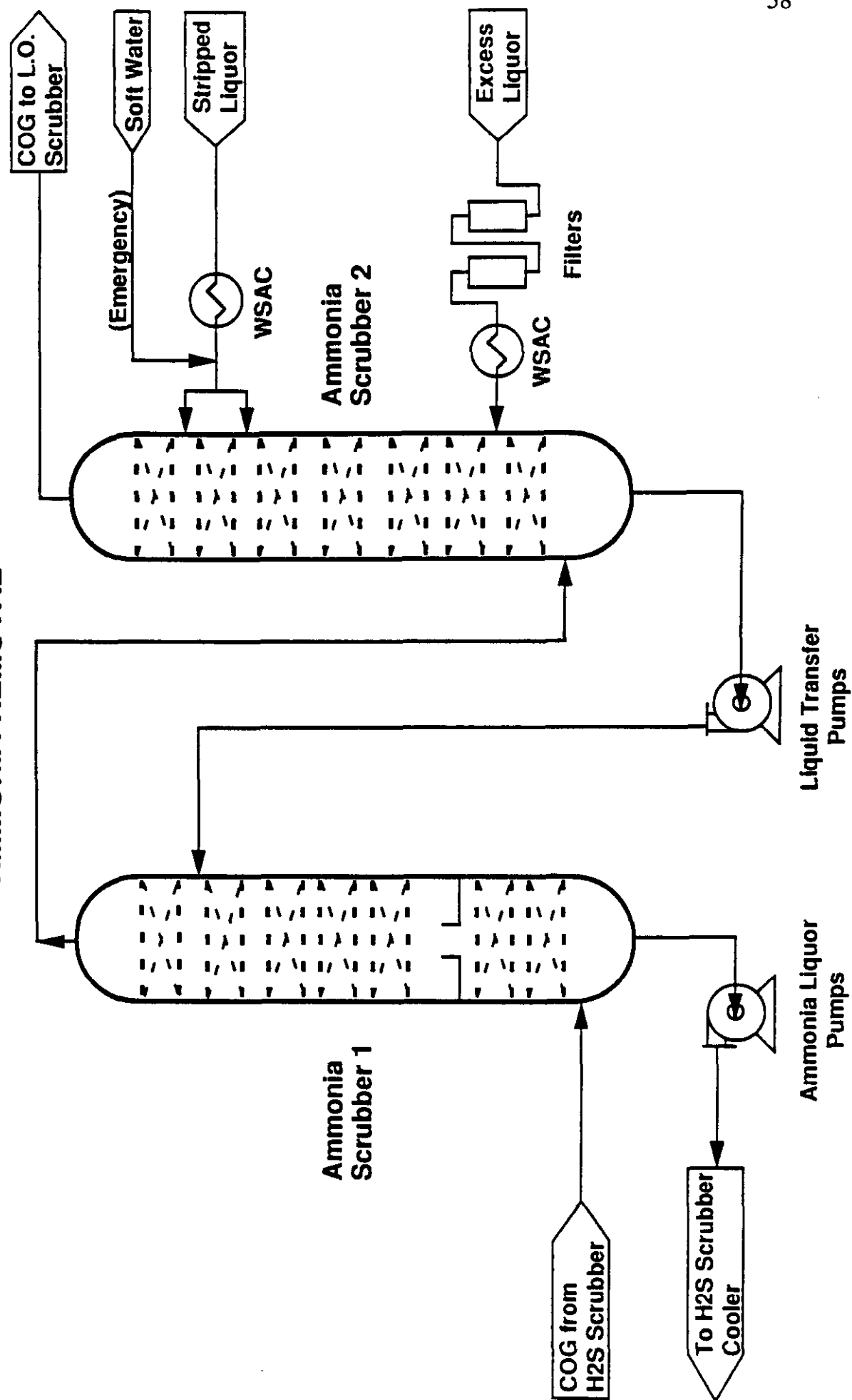
### 3.3.2 Capacity

The unit can operate over a wide range of flow rates and NH<sub>3</sub> concentrations. The unit will give satisfactory performance at 80 to 100 percent of the design flowrates with NH<sub>3</sub> concentrations up to 300 Gr NH<sub>3</sub>/100 SCF.

### 3.3.3 Schematic Flow Diagram

A schematic flow diagram of the Ammonia Removal stage is shown on Figure 3.3.3-1.

Figure 3.3.3-1

**AMMONIA REMOVAL**

### 3.3.4 Major Equipment List and Spares

<u>Equipment No.</u>	<u>Description/Size</u>	<u>Operating</u>	<u>Spare</u>
62.131.1-2	Gravel Filters 8' Dia. x 15' Ht.	1	1
67.111	Ammonia Scrubber 1 16'-6" Dia. x 107'-3" Ht., CS	1	
67.112	Ammonia Scrubber 2 16'-6" Dia. x 103'-11" Ht., CS	1	
67.262.1-2	Ammonia Liquor Pumps 350 GPM, 70 PSIG, 1780 RPM, CI	1	1
67.362.1-2	Liquor Transfer Pumps 350 GPM, 70 PSIG, 1780 RPM, CI	1	1

### 3.3.5 Mass Balance

The mass balance and operating conditions for the Ammonia Scrubber section of the plant are given on Process Flow Diagram No. 192875 located in Section 3.9.

### 3.3.6 Utility Requirements

Utilities required for NH<sub>3</sub> Removal are listed below. Overall utility requirements for the COG cleaning system, excluding Ammonia Destruction and Claus Sulfur Recovery are shown in Section 3.8.

- o Electricity
- o Instrument air
- o Low pressure steam (clean out)
- o Industrial water (service)
- o Medium pressure steam (service)
- o Plant air (service)

### 3.3.7 Catalyst and Chemical Requirements

No purchased chemicals are required. Reagents for NH<sub>3</sub> removal are generated within the coke making process, or in the COG cleaning system. Chemicals required for absorbing NH<sub>3</sub> from the COG are the following:

<u>Chemical</u>	<u>Use</u>	<u>Source</u>
Excess Ammonia Liquor	Ammonia Absorption	Coking Process
Stripped Liquor	Ammonia Absorption	COG Cleaning
Flushing Liquor (Emergency)	Direct Contact Gas Cooling	Coking Process
Tar (Emergency)	Dissolve Naphthalene	Coking Process
Soft water (Emergency)	Ammonia Absorption	Boiler House

### 3.3.8 Design Considerations

**COG Flow** - This flow is one of the factors used to determine vessel size.

**Temperature** - Operating temperatures effect the component equilibriums between the vapor and the liquid phases. Lower operating temperatures increase  $\text{NH}_3$  absorption.

**$\text{NH}_3$  and  $\text{CO}_2$  Concentration** -  $\text{NH}_3$  and  $\text{CO}_2$  concentrations effect the scrubber efficiency. High  $\text{NH}_3$  and low  $\text{CO}_2$  concentrations are desired.

**On-Stream Factor** - Plant on-stream factor is related to equipment reliability. To maintain a high on-stream factor, the  $\text{NH}_3$  Scrubbers may be operated individually instead of in series.

**Clean-out** - The unit should be designed to eliminate fouling and for mechanical reliability. Each time a piece of equipment is taken out of service, some pollution occurs from draining, flushing, and cleaning.

**Cooling Equipment** - Open loop cooling systems allow contaminants such as  $\text{NH}_3$ ,  $\text{H}_2\text{S}$ , and benzene to escape to the atmosphere. Use of a closed loop cooling system eliminates this problem.

**Pressure Drop** - To minimize the Booster Compressor Hp requirements, the absorption system must operate at a low pressure drop. This is done by proper line sizing, packing selection, and equipment location.

**Contact Time** - Adequate gas/liquid contact time must be allowed to absorb the  $\text{NH}_3$ . Scrubber length and diameter are selected to provide the required contact time.

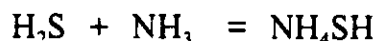
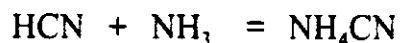
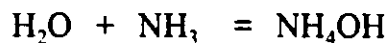
**Materials of Construction** - COG contains a variety of chemicals which have the potential to be corrosive under certain environments. Material selection was based on stream composition and operating conditions.

### 3.3.9 Normal Operation

Reference: Process Flow Diagram No. 192875 in Section 3.9.

COG leaves the top of the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler and flows to the bottom of Ammonia Scrubber 1. Scrubbing fluid is stripped liquor from the Free-Ammonia Still. A portion of this liquor is cooled in one of the WSAC coils, split into Streams 3 and 4, and then both streams are charged to the top of Ammonia Scrubber 2. The liquid descends the packed scrubber and contacts the COG from Ammonia Scrubber 1. The remainder of the liquid used in Ammonia Scrubber 2, Stream 5, is excess liquor from storage that is cooled and charged to the bottom of the column. Total liquor from the bottom of Ammonia Scrubber 2 is pumped to the top of Ammonia Scrubber 1 by the Liquor Transfer Pumps. The liquor descends the column and contacts the COG received from the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler. Clean gas, Stream 2, containing less than 55 Gr  $\text{H}_2\text{S}$  and 4 Gr  $\text{NH}_3$ /100 SCF leaves the top of Ammonia Scrubber 2 and flows to the light oil scrubbers outside the gas cleaning plant's limits. Rich- $\text{NH}_3$  liquor, Stream 7, is pumped by the Ammonia Liquor Pumps to the top of the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler.

The relevant reactions in the Ammonia Scrubbers are:



### 3.3.10 Start-up/Upset/Shutdown Conditions

Under upset conditions, processing variables can be adjusted to compensate for the upsets. Variables that can be changed are:

Liquid Temperatures - Lower operating temperatures increase  $\text{NH}_3$  absorption.

Liquid flow rates - Higher liquid flow rates favor  $\text{NH}_3$  absorption.

In addition, if the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler is out of service, cooling may be performed in Ammonia Scrubber 1 by setting appropriate valves.

### 3.3.11 Environmental Considerations

Gaseous emissions from gas cooling - Use of closed loop indirect coolers eliminates the emissions problems associated with direct gas quenching.

Liquid contamination from cleaning equipment - Equipment that has a history of fouling must be opened for cleaning and flushing. Use of non-fouling packing and tar addition to dissolve naphthalene eliminates this problem.

Absorbent - Absorbent regenerability, by-product formation and waste disposal were considered. Since water is the absorbent, no new problems were introduced beyond those already existing in treating COG bleed streams.

Water use and contamination - It is important to reduce make-up water use and prevent its contamination. This processing step does not require water make-up except in the WSAC.

Outside chemical use and disposal - The process should avoid the use of chemicals which must be disposed or produce a disposable by-product. No new problems were introduced beyond those already in the existing COG treating system.

### 3.4 DEACIFICATION

#### 3.4.1 Process Description

Reference: Schematic flow diagram in Section 3.4.3.

The purpose of this unit is to recover the acid gases from the rich liquor. This is accomplished by stripping with steam and  $\text{NH}_3$  vapors.

Enriched- $\text{NH}_3$  liquor from the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler is pumped through two sets each of two series/parallel-connected plate heat exchangers. In these exchangers, the enriched liquor is preheated by exchange, first with deacidified water, then with Free-Ammonia Still effluent up to  $167^\circ\text{F}$  before entering the top of the Deacidifier scrubber. This is a trayed unit in which the acid gases are stripped from the descending liquor by means of a rising flow of  $\text{NH}_3$  vapor from the Free-Ammonia Still and steam. The hot,  $\text{NH}_3$  rich deacidified water leaving the base of the Deacidifier, is pumped through the previously mentioned heat exchangers and a coil of the WSAC and is returned to the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler for another absorption cycle. In addition to the duty Deacidifier, a standby Deacidifier is provided.

The overhead vapors leaving the top of the Deacidifier are a mixture of  $\text{NH}_3$ , steam and the acid gases  $\text{H}_2\text{S}$ ,  $\text{HCN}$  and  $\text{CO}_2$  at a temperature of approximately  $187^\circ\text{F}$ . These gases are cooled to approximately a temperature of  $165^\circ\text{F}$  in a partial condenser, and the resulting liquid condensate is returned to the top section of the Deacidifier. The partially cooled acid gas/ $\text{NH}_3$  vapor stream, having a much reduced water vapor content, is passed to the Ammonia Destruction and Claus Sulfur Recovery units.

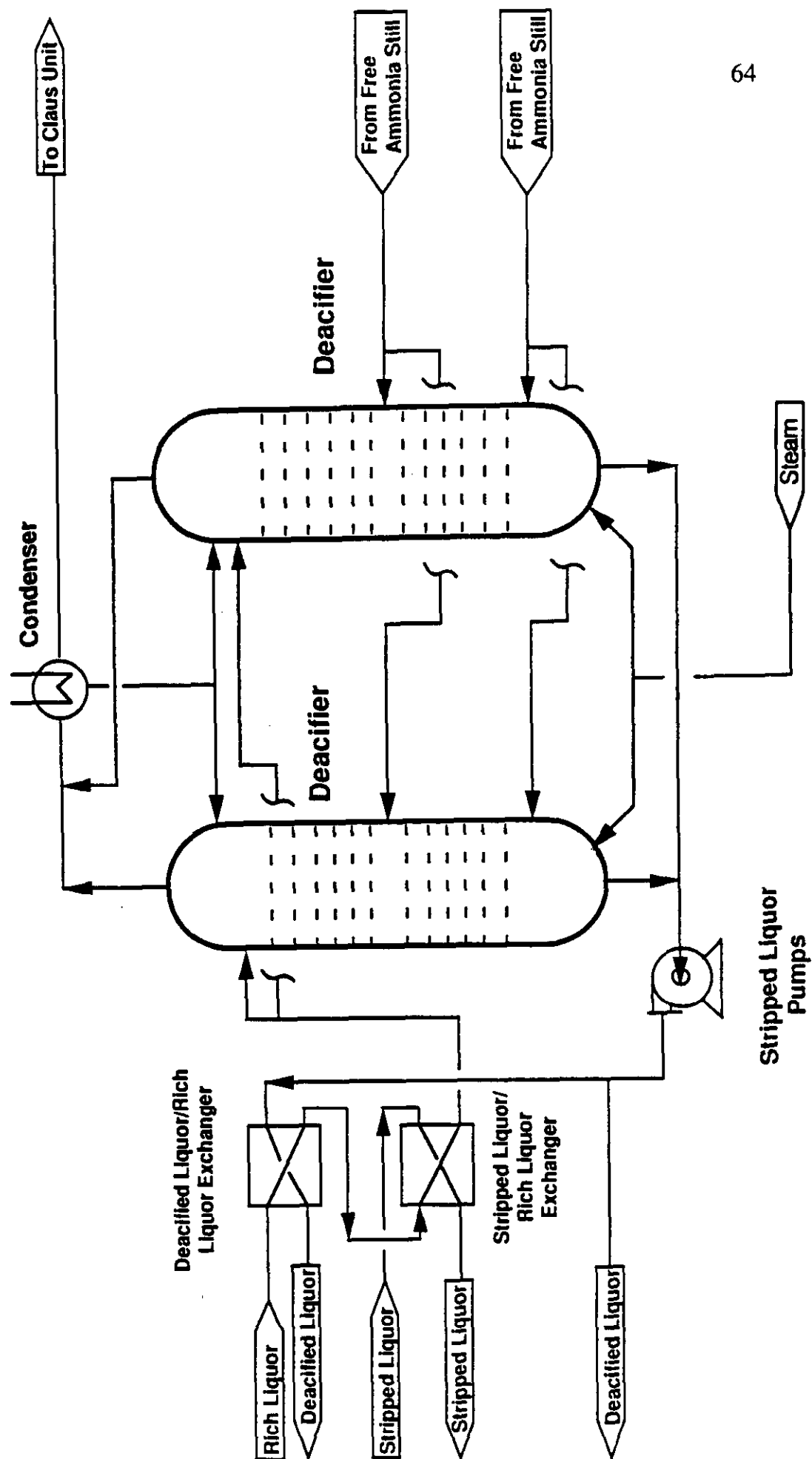
#### 3.4.2 Capacity

Each Deacidifier Column has the capacity to deacidify the design rate flow of rich liquor from the  $\text{H}_2\text{S}$  Scrubber.

#### 3.4.3 Schematic Flow Diagram

A schematic flow diagram of the Deacidification stage is shown on Figure 3.4.3-1.

**Figure 3.4.3-1**  
**DEACIFIERS**



### 3.4.4 Major Equipment List and Spares

<u>Equipment No.</u>	<u>Description/Size</u>	<u>Operating</u>	<u>Spare</u>
68.121.1-2	Stripped Liquor/Rich Liquor Heat Exchangers 430 Ft <sup>2</sup>	1	
68.168.1-2	Salt Water Booster Pumps 350 GPM	1	1
68.511.1-2	Deacifiers 8' Dia. x 37' Ht., Hastelloy C4	1	1
68.522	Condenser 406 Ft <sup>2</sup> , Graphite CS/PTFE	1	
68.622.1-2	Deacidified Liquor/Rich Liquor Exchanger 4860 Ft <sup>2</sup> , Titanium	1	
68.761.1-2	Deacidifier Pumps 880 GPM, 80 PSIG, 1780 RPM, Hastelloy C4	1	1

### 3.4.5 Mass Balance

The mass balance and operating conditions for the Deacidification section are given on Process Flow Diagram No. 192903 located in Section 3.9.

### 3.4.6 Utility Requirements

Utilities required for the Deacidifiers are listed below. Overall utility requirements for the COG cleaning system, excluding Ammonia Destruction and Claus sulfur recovery, are shown in Section 3.8.

- o Cooling water (Salt water)
- o Electricity
- o Instrument air
- o Medium pressure steam (service)
- o Industrial water
- o Plant air (service)
- o Softened water

#### 3.4.7 Catalyst and Chemical Requirements

No purchased chemicals are required.

#### 3.4.8 Design Considerations

**Temperature** - Operating temperature affects the equilibrium between the vapor and liquid. The temperature is adjusted to balance operating costs and acid gas removal.

**Pressure** - Pressure is a consideration for acid gas removal. Low pressure favors increased acid gas removal.

**Tower Packing** - A packing must be selected that has low fouling tendency, low pressure drop, and provides good contact between vapor and liquid.

**Equipment Reliability** - A spare Deacidifier is installed to avoid shutdown for column or mechanical equipment problems.

**Flow Rates** - Flow rates are important because they determine the scrubber size, scrubber temperature, and scrubber performance with respect to acid gas removal.

**Cooling Equipment** - Proper design is required to cool the gas, condense water, and separate the water from the vapor.

**Materials of Construction** - COG contains a variety of chemicals which have the potential to be corrosive under certain environments. Material selection was based on stream composition and operating conditions.

#### 3.4.9 Normal Operation

Reference: Process Flow Diagram No. 192903 in Section 3.9.

Rich Liquor from the H<sub>2</sub>S Scrubber, Stream 1, is heated from a temperature of 86°F to 153°F by interchanging heat with the deacidified liquor and stripped liquor. This stream, containing the acid gases and NH<sub>3</sub>, enters the top of the trayed Deacidifier and descends through the column. Medium pressure steam, as required from the steam header,

Stream 11, and  $\text{NH}_3$  from the Free-Ammonia Still contact the descending liquor and strip the acid gases. Overhead vapor at a temperature of  $185^\circ\text{F}$  is cooled in the condenser to condense water. Gas from the condenser, Stream 12, is sent to the Ammonia Destruction Unit. A portion of the hot  $\text{NH}_3$ -rich bottoms is pumped to the Free-Ammonia Still. The remaining liquor, Stream 2, is cooled by interchange and by WSAC and returned to the  $\text{H}_2\text{S}$  Scrubber/Secondary Cooler.

#### 3.4.10 Start-up/Upset/Shutdown Conditions

A spare Deacifier Scrubber is provided to handle upset conditions.

#### 3.4.11 Environmental Considerations

A spare Deacifier is installed to insure the process stays on line. Failure of this unit would cause a plant shutdown.

### 3.5 AMMONIA STILL

#### 3.5.1 Process Description

Reference: Schematic flow diagram in Section 3.5.3.

Two Free-Ammonia Stills are used to remove the free and combined  $\text{NH}_3$  from the absorption liquor. The Free-Ammonia Stills are packed and trayed scrubbers in which the  $\text{NH}_3$ -rich deacidified liquor is stripped of free  $\text{NH}_3$  and residual acid gas components. Low-pressure steam is used as the stripping medium, including steam produced in the Claus Sulfur Recovery Unit.  $\text{NH}_3$ -rich vapors are passed from the appropriate intermediate and top sections of the Free-Ammonia Still into the optimum locations in the Deacidifier to produce a strong  $\text{NH}_3$ -rich deacidified liquor for  $\text{H}_2\text{S}$  scrubbing.

The hot stripped liquor leaving the base of the Free-Ammonia Still is divided into two streams. To maintain the overall plant water balance, a bleed stream is passed to the Fixed-Ammonia Still. The remaining Free-Ammonia Still effluent is partially cooled in two exchangers preheating the Deacidifier feed liquor, and further cooled in a coil of the WSAC before passing to the Ammonia Scrubber as an absorbing liquor. The Fixed Ammonia Still is a trayed scrubber. In the upper section, the feed liquor is intensively stripped of any remaining free acid gases.

Beneath the feed tray, caustic soda solution is added to release  $\text{NH}_3$ , which is chemically bound as fixed salts, e.g. chlorides.

Low-pressure steam is used as the stripping medium, and the still head vapors are passed to an intermediate location on the Free-Ammonia Still. Using this system, the required low  $\text{NH}_3$  concentration in the effluent is achieved. The stripped effluent from the Fixed-Ammonia Still is cooled and forwarded to the existing effluent treatment plant.

A standby Free-Ammonia Still is provided. This still is capable of performing either the Free or Fixed-Ammonia Still duties.

### 3.5.2 Capacity

The Ammonia Stills are designed to process the design flow rate of Deacidifier liquor containing the design concentrations of  $\text{NH}_3$ . To maintain this capacity, a spare Free-Ammonia Still is installed. Also, if necessary, either one of the Free-Ammonia Stills may be used as a Fixed- Ammonia Still.

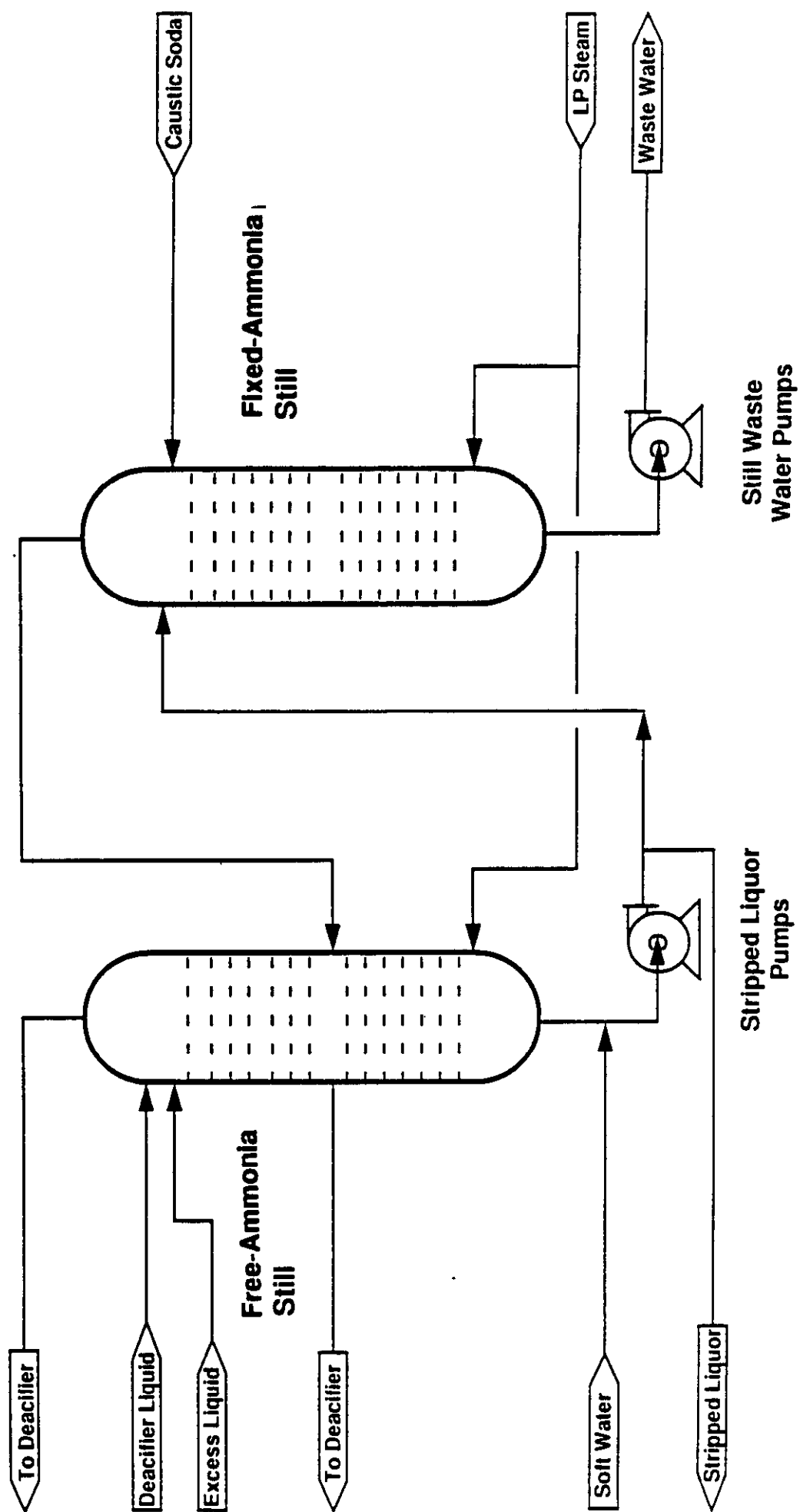
### 3.5.3 Schematic Flow Diagram

A schematic flow diagram of the Ammonia Still stage is shown on Figure 3.5.3-1.

### 3.5.4 Major Equipment List and Spares

<u>Equipment No.</u>	<u>Description/Size</u>	<u>Operating</u>	<u>Spare</u>
68.112.1-2	Free Ammonia Still 8' Dia. x 40' Ht., CI	1	1
68.114	Fixed Ammonia Still 6' Dia. x 36'-3" Ht. , CI	1	
68.163.1-2	Stripped Liquor Pumps 530 GPM, 79 PSIG, 1800 RPM	1	1
68.165.1-2	Still Waste Water Pumps 310 GPM, 51 PSIG, 1780 RPM	1	1

**Figure 3.5.3-1**  
**AMMONIA STILLS**



### 3.5.5 Mass Balance

The mass balance and operating conditions for the Ammonia Stills are shown on Process Flow Diagram No. 192903 located in Section 3.9.

### 3.5.6 Utility Requirements

Utilities required for the Ammonia Stills are listed below. Overall utility requirements for the coke oven gas cleaning system, excluding Ammonia Destruction and Claus Sulfur Recovery are shown in Section 3.8.

- o Softened water
- o Low Pressure steam
- o Plant air (Service)
- o Industrial water
- o Medium pressure steam
- o Soft water

### 3.5.7 Catalyst and Chemical Requirements

A fraction of the  $\text{NH}_3$  in the deacidified liquor is chemically combined, and cannot be removed by simple steam stripping. To release this  $\text{NH}_3$ , caustic soda is added to the top section of the Free-Ammonia Still.

### 3.5.8 Design Considerations

Temperature - Temperature sets the relationship between the component compositions in the gas and liquid.

Ammonia Concentrations - The  $\text{NH}_3$  is present as free and combined. The quantity of fixed  $\text{NH}_3$  sets the caustic soda requirements.

Pressure - Stripping of  $\text{NH}_3$  is favored by low pressure operation.

Liquid Flow Rates - In part, the Deacidifier liquid flow rate sets the caustic soda and steam rates.

Scrubber Internals - The scrubbers must be designed for good liquid and caustic soda contact and for efficient  $\text{NH}_3$  stripping.

**Pressure Drop** - Pressure drop in the scrubbers is not as critical as it is in the H<sub>2</sub>S Scrubber/Secondary Cooler and Ammonia Scrubber, however it should be as low as practical.

**Equipment Reliability** - A spare Free-Ammonia Still is provided. Also, if required, either one of the Free-Ammonia Stills may be used as a spare for the Fixed-Ammonia Still.

**Materials of Construction** - COG contains a variety of chemicals which have the potential to be corrosive under certain environments. Material selection was based on stream composition and operating conditions.

### 3.5.9 Normal Operation

**Reference:** Process Flow Diagram No. 192903 in Section 3.8.

Deacidified liquor is fed to the top tray of the Free-Ammonia Still. This liquor descends through the scrubber packing and trays and is contacted with NH<sub>3</sub> and steam vapors from the Fixed-Ammonia Still, and low pressure steam, Stream 10, from the Claus unit and exhausters. NH<sub>3</sub> and steam vapors exit at the top of the still and enter the Deacidifier Scrubber. Stripped liquor leaves the bottom of the still, and a slipstream is pumped to the top of the Fixed-Ammonia Still. The liquor descends the scrubber. Beneath the feed tray, caustic soda is added to free the fixed-NH<sub>3</sub>. Low pressure steam is charged at the bottom of the scrubber, ascends the scrubber, and sweeps the NH<sub>3</sub> overhead and into the middle section of the Free-Ammonia Still. Waste water is removed from the bottom of the Fixed-Ammonia Still, pumped through salt water coolers and discharged to the bio-oxidation plant, Stream 6.

### 3.5.10 Start-up/Upset/Shutdown Conditions

A spare Free-Ammonia Still is provided to handle upset conditions. Also, if required, either one of the Free-Ammonia Stills may be used as a spare for the Fixed-Ammonia Still.

### 3.5.11 Environmental Considerations

**Air Pollution** - The system is relatively simple with few rotating pieces of equipment. The system should be designed with a minimum number of openings and reliable pumps should be selected.

Water Pollution - Steam usage should be kept to a minimum because it will eventually be contaminated when it is condensed. Also, the pressure of the salt water coolant used to cool the still waste water should be maintained above the process pressure so that process fluid can not leak into the sea water.

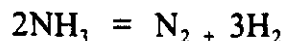
### 3.6 AMMONIA DESTRUCTION

#### 3.6.1 Process Description

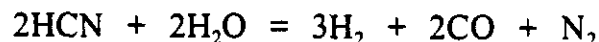
Reference: Schematic flow diagram in Section 3.6.3.

The partially-cooled overhead vapors from the Deacidifier Scrubber contain all the  $\text{H}_2\text{S}$  and  $\text{NH}_3$  removed from the COG together with HCN and  $\text{CO}_2$ . This acid gas mixture is mixed with air and fed to a specially designed burner at the top of the catalytic reactor. The reactor contains a nickel catalyst which is insensitive to the presence of sulfur compounds, and operates at about  $2000^\circ\text{F}$ . Within the catalytic reactor,  $\text{NH}_3$ , HCN, and organic carbon compounds are completely decomposed to produce  $\text{CO}$ ,  $\text{CO}_2$ ,  $\text{N}_2$ ,  $\text{H}_2$ , and water. In the presence of water vapor, carbon is not deposited on the catalyst surface at these temperatures.

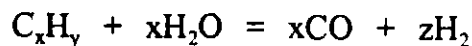
Key reactions occurring include the following:



HCN decomposition:

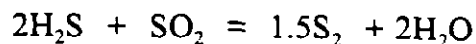
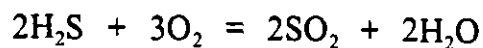


Tar decomposition:



Following the catalytic decomposition, additional air is added to the process gases to provide the stoichiometric ratio of  $\text{O}_2$  to  $\text{H}_2\text{S}$  for the Claus reaction to proceed.

Claus reaction:



### 3.6.2 Capacity

The Ammonia Destruction Reactor has the capacity to destroy the design flows of  $\text{NH}_3$ , HCN, and hydrocarbons.

### 3.6.3 Schematic Flow Diagram

A schematic flow diagram of the Ammonia Destruction stage is shown on Figure 3.6.3-1.

### 3.6.4 Major Equipment List and Spares

<u>Equipment No.</u>	<u>Description/Size</u>	<u>Operating</u>	<u>Spare</u>
92.954	Catalytic Reactor 11'-2" Dia. x 32'-2' Ht., CS	1	
92.963.1-2	Reaction Air Blowers 149,000 SCFH, 18.9 PSIG, CS	1	1

### 3.6.5 Mass Balance

The mass balance and operating conditions for the Ammonia Destruction Unit are shown on Process Flow Diagram No. 193476 located in Section 3.9.

### 3.6.6 Utility Requirements

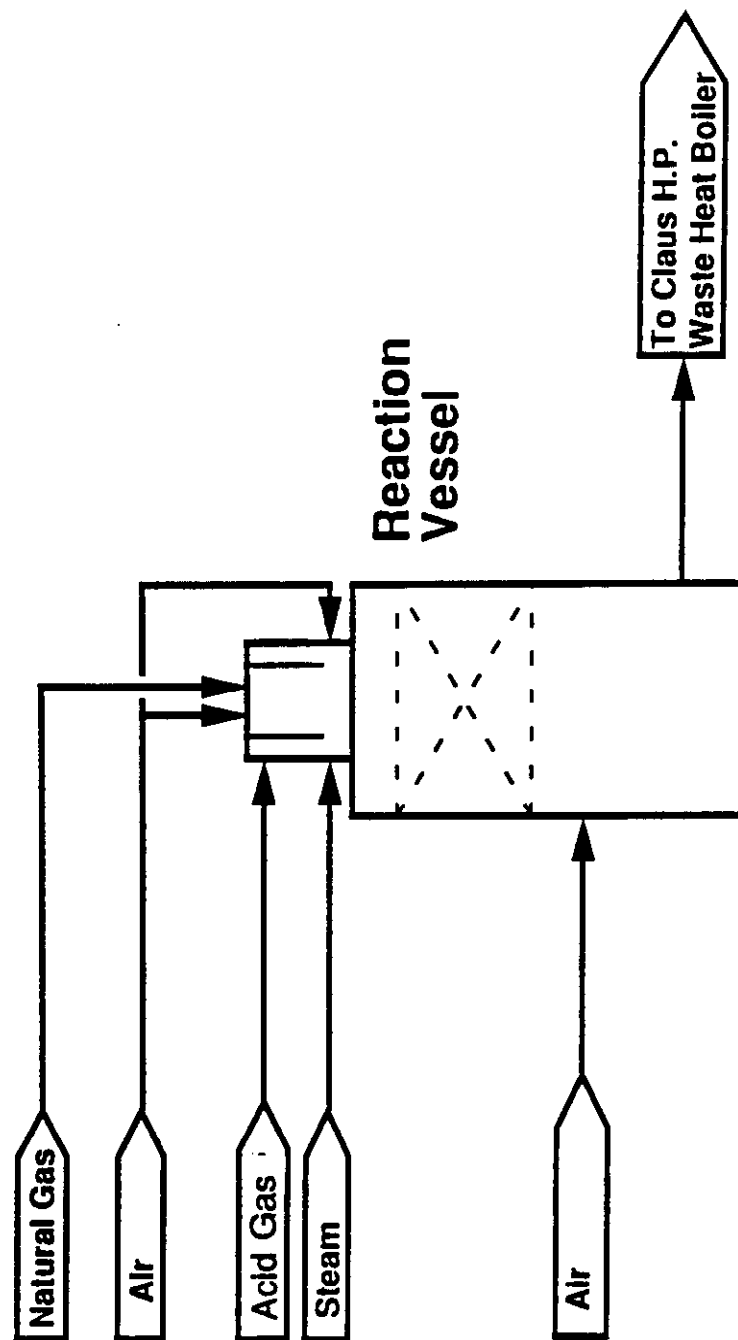
Utilities required for Ammonia Destruction are listed below. Overall utility requirements for the Ammonia Destruction and Claus Sulfur Recovery Unit are shown in Section 3.8.

- o Natural gas
- o Plant air
- o Electricity
- o Instrument air

### 3.6.7 Catalyst and Chemical Requirements

A sulfur resistant catalyst containing nickel is used to decompose the  $\text{NH}_3$ , HCN, and hydrocarbons.

**Figure 3.6.3-1**  
**AMMONIA DESTRUCTION**



### 3.6.8 Design Considerations

**Temperature** - The reactor must be operated at a high temperature to insure complete decomposition of the  $\text{NH}_3$ ,  $\text{HCN}$ , and hydrocarbons.

**Catalyst Type** - The catalyst must be thermally stable, reactive at high temperature, and resistant to sulfur poisoning.

**Start-up** - The catalyst must be heated to operating temperature before gas is charged. This is accomplished by burning natural gas in a startup burner.

**Refractory** - The refractory must be thermally stable, have good insulating properties and be resistant to attack by the process gas.

**Coke Formation** - When hydrocarbons are decomposed, coke is formed. This can be eliminated by injecting steam into the process stream. In the presence of steam the hydrocarbons are converted to  $\text{H}_2$  and  $\text{CO}$ .

**Air Requirements** - Air is injected to convert one third of the  $\text{H}_2\text{S}$  to  $\text{SO}_2$  in order to produce the required amount of  $\text{SO}_2$  for the Claus reaction. This air flow is monitored closely. Insufficient air will allow excess  $\text{H}_2\text{S}$  to be recycled in the Claus tail gas to the Primary Coolers. Excess air will allow excess  $\text{SO}_2$  to be recycled in the Claus tail gas to the Primary Coolers.

### 3.6.9 Normal Operation

Reference: Process Flow Diagram No.193476 in Section 3.9.

Cool Deacidified gas, Stream 1, enters the reactor beneath the start-up burner. Steam, air and natural gas can also be fed into this area as required. During normal operation, air and fuel are fed to control the catalyst bed temperature, and steam is added to control coke formation. Within the reactor, which operates at  $2000^\circ\text{F}$ ,  $\text{NH}_3$ ,  $\text{HCN}$  and hydrocarbons are decomposed. Gas at the outlet of the catalyst bed is contacted with additional air to convert one-third of the  $\text{H}_2\text{S}$  to  $\text{SO}_2$ , in order to produce the  $\text{H}_2\text{S}/\text{SO}_2$  ratio required for the Claus reaction. All process air is supplied by the Combustion Air Blower.

### 3.6.10 Start-up/Upset/Shutdown Conditions

A natural gas fired startup burner is provided to preheat the catalyst bed to the required operating temperature.

The reactor control system is capable of handling upset and shutdown conditions. The control system monitors the operation, adjusts fuel, air and steam flows, and may shut the reactor down under certain upset conditions.

### 3.6.11 Environmental Considerations

Air Pollution - The system should have a minimum number of openings to prevent gas leaks.

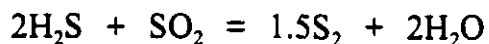
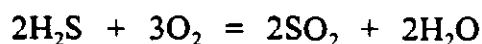
## 3.7 CLAUS SULFUR RECOVERY UNIT

### 3.7.1 Process Description

Reference: Schematic flow diagram in Section 3.7.3.

The first stage of the Claus reaction, which is thermally initiated, takes place at the outlet of the Ammonia Destruction Unit. At the outlet, additional air is added to the process gases to provide the stoichiometric ratio of SO<sub>2</sub> to H<sub>2</sub>S for the Claus reaction to proceed.

Claus reaction:



About 65 to 75 percent of the sulfur recovered from the COG is produced at this point and is condensed as the process gases are cooled to a temperature of approximately 475°F in the High Pressure (HP) Waste Heat Boiler. This boiler generates 565 PSIG steam from deaerated and chemically treated boiler feedwater.

Gas from the HP Waste Heat Boiler is split into two streams. One stream passes through the Low Pressure (LP) Waste Heat Boiler where the gas is further cooled by generating low pressure steam. The other stream by-passes the LP Boiler and rejoins the cooled gas stream exiting the LP Boiler. The by-pass flow is controlled to regulate the temperature of the gas which is then fed to the First Stage Claus Reactor. The bulk of the remaining  $H_2S$  is converted to sulfur in this catalytic reactor.

Gas from the First Stage Claus Reactor is cooled in one side of the Sulfur Condenser where low pressure steam is generated. The cooled first stage gas from the Sulfur Condenser is then reheated by medium pressure steam in the Process Gas Heater, and fed to the Second Stage Claus Reactor where the remaining  $H_2S$  is converted to  $SO_2$ . Effluent from this reactor passes through the second tube side of the Sulfur Condenser to condense sulfur and generate additional low pressure steam. The condensed sulfur is collected in the Sulfur Pit.

The hot tail gas from the Sulfur Condenser is recycled back to the COG main upstream from the Coal Chemical Plant Plant B primary coolers. Recycling the tail gas in this manner maintains the overall  $H_2S$  removal at the design value.

### 3.7.2 Capacity

The HP Waste Heat Boiler, the LP Waste Heat Boiler, the Sulfur Condenser and the Claus Reactors have the capacity to convert  $H_2S$  and condense sulfur at the design COG flow rates.

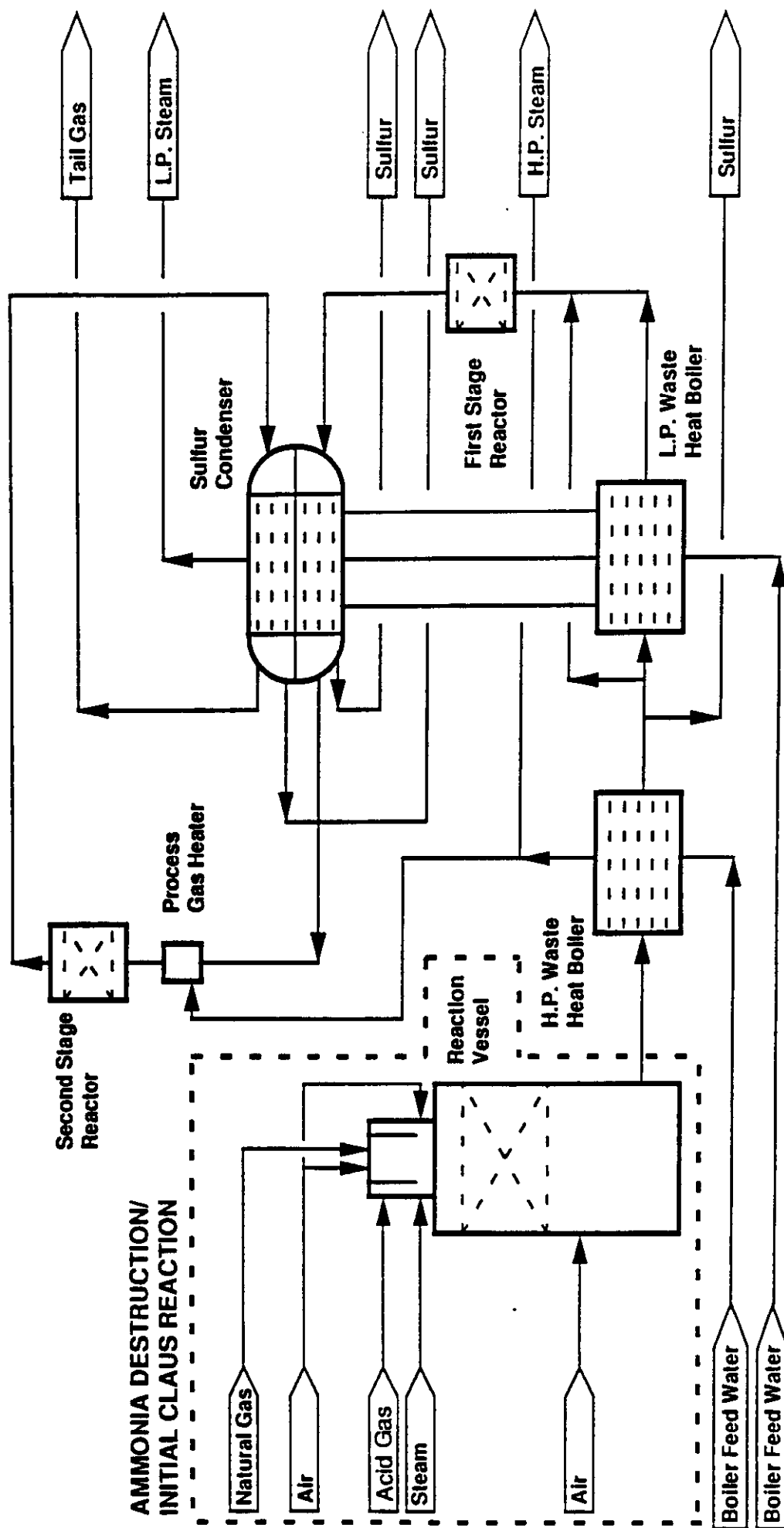
### 3.7.3 Schematic Flow Diagram

A schematic flow diagram of the Claus Sulfur Recovery Unit is shown on Figure 3.7.3-1.

### 3.7.4 Major Equipment List and Spares

<u>Equipment No.</u>	<u>Description/Size</u>	<u>Operating</u>	<u>Spare</u>
68.928	Stand-by Thermal Oxidizer 75,500 SCFH Dry/110,800 SCFH Wet	1	
92.111	First Stage Claus Reactor 11'-2" Dia. x 32' Ht.	1	

**Figure 3.7.3-1**  
**CLAUS SULFUR RECOVERY**



<u>Equipment No.</u>	<u>Description/Size</u>	<u>Operating</u>	<u>Spare</u>
92.113	Second Stage Claus Reactor 11'-2" Dia. x 32' Ht.	1	
92.124	Sulfur Condenser 50 PSIG Steam/2.7 PSIG Process Gas	1	
92.125	Process Gas Heater 565 PSIG Steam/4.2 PSIG Process Gas	1	
92.921	H. P. Waste Heat Boiler 15,950 lb/hr Steam/249,400 SCFH Process Gas	1	
92.925	L. P. Waste Heat Boiler 11,900 lb/hr Steam/186,200 SCFH Process Gas	1	

### 3.7.5 Mass Balance

The mass balance and the operating conditions for the Claus Sulfur Recovery Unit are shown on Process Flow Diagram No. 193476 located in Section 3.9.

### 3.7.6 Utility Requirements

Utilities required for the Claus Sulfur Recovery Unit are listed below. Overall utility requirements for the Ammonia Destruction and Claus Sulfur Recovery Unit are shown in Section 3.8.

- o Boiler feed water
- o Industrial water
- o Medium pressure steam
- o Process air
- o Nitrogen
- o Electricity
- o Instrument air
- o City water

### 3.7.7 Catalyst and Chemical Requirements

The following catalysts and chemicals are required for the Claus Sulfur Recovery Unit:

- o Claus Reactor Catalyst
- o Boiler feed water chemicals

### 3.7.8 Design Considerations

**Catalyst** - The catalyst must have long life and be capable of reacting  $\text{H}_2\text{S}$  and  $\text{SO}_2$  to produce elemental sulfur.

**Temperature** - The temperature must be high enough to have adequate reaction rate with consideration given to catalyst stability.

**Pressure** - Pressure is kept to a minimum from vessel design and process considerations.

**Heat Duties** - Consideration must be given to how to extract the heat of reaction from the Claus reactors and control the feed temperature to the second Claus Reactor.

**Mass Balance** - It is important to control the air injected and the quantity of  $\text{SO}_2$  formed.

**Layout** - Consideration must be given to equipment layout for heat recovery, sulfur collection and pressure drops.

**Refractory** - The refractory must be thermally stable, have good insulating properties and be resistant to attack by the process gas.

**Materials of Construction** - Coke oven gas contains a variety of chemicals which have the potential to be corrosive under certain environments. Material selection was based on stream composition and operating conditions.

### 3.7.9 Normal Operation

**Reference:** Process Flow Diagram No. 193476 in Section 3.8.

In the bottom section of the catalytic reactor, air is injected into the hot  $\text{NH}_3$ -free gas to convert one third of the  $\text{H}_2\text{S}$  to  $\text{SO}_2$  in order to provide the appropriate ratio of  $\text{SO}_2$  to  $\text{H}_2\text{S}$  required for the Claus reaction. Air is supplied by the Combustion Air Blowers. The reacted gas is analyzed, on stream, for  $\text{H}_2\text{S}$  and  $\text{SO}_2$  content and the analysis is used to set the air flow rate, Stream 2. In this reactor, the Claus reaction then proceeds to greater than 65 percent completion. The hot gas exits the reactor and enters the HP Waste Heat Boiler where sulfur is condensed and high pressure steam is generated. Make-up water to the boiler, Stream 16, is deaerated

and treated chemically. The process gas is further cooled in the LP Waste Heat Boiler where low pressure steam is generated.

Feed to the First Stage Claus Reactor is a combination of hot gas from the HP Waste Heat Boiler and cooled gas from the LP Waste Heat Boiler. The by-pass rate is regulated to control the gas mixture temperature, Stream 7. The Claus reaction is nearly completed in the First Stage Claus Reactor. This reaction is exothermic with the temperature increasing from a temperature of 473°F at the inlet to a temperature of 541°F at the outlet, Stream 8. Sulfur vapor in this stream is condensed in one side of the sulfur condenser. The cool gas from the condenser is reheated in the Process Gas Heater by condensing medium pressure steam. The reheated gas, Stream 9, enters the Second Stage Claus Reactor at a temperature of 410°F and exits at a temperature of 442°F, Stream 10. This gas is also cooled in the Sulfur Condenser. The hot tail gas exits the condenser at a temperature of 228°F, Stream 11, and is recycled back to the COG main upstream of the Primary Coolers.

Sulfur from each condensing stage is collected in the Sulfur Pit.

#### 3.7.10 Startup/Upset/Shutdown Conditions

For start-up purposes, a heater is provided to preheat the catalyst beds and avoid sulfur condensation.

An emergency thermal oxidizer and stack are provided for periods when the Ammonia Destruction and Claus Sulfur Recovery units are not in operation.

#### 3.7.11 Environmental Considerations

Air Pollution - The unit must be carefully constructed and sealed to prevent escape of  $H_2S$  and  $SO_2$ .

### 3.8 UTILITIES

#### 3.8.1 $H_2S$ and Ammonia Removal and Recovery

The utility consumptions identified in this section are the total requirements for secondary cooling,  $H_2S$  and  $NH_3$  scrubbing,  $NH_3$  distillation, and deacidification.

<u>Utility</u>	<u>Usage</u>
Electric Power Absorbed, KW/hr	626
LP Steam, lb/hr	44,065
Industrial Water, GPH	27,500
Salt water, GPH	75,420
City water, GPH	
Normal	0
Maximum	6,000

### 3.8.2 Ammonia Destruction and Claus Sulfur Recovery

The utility consumptions identified in this section are the total requirements for ammonia destruction and sulfur recovery.

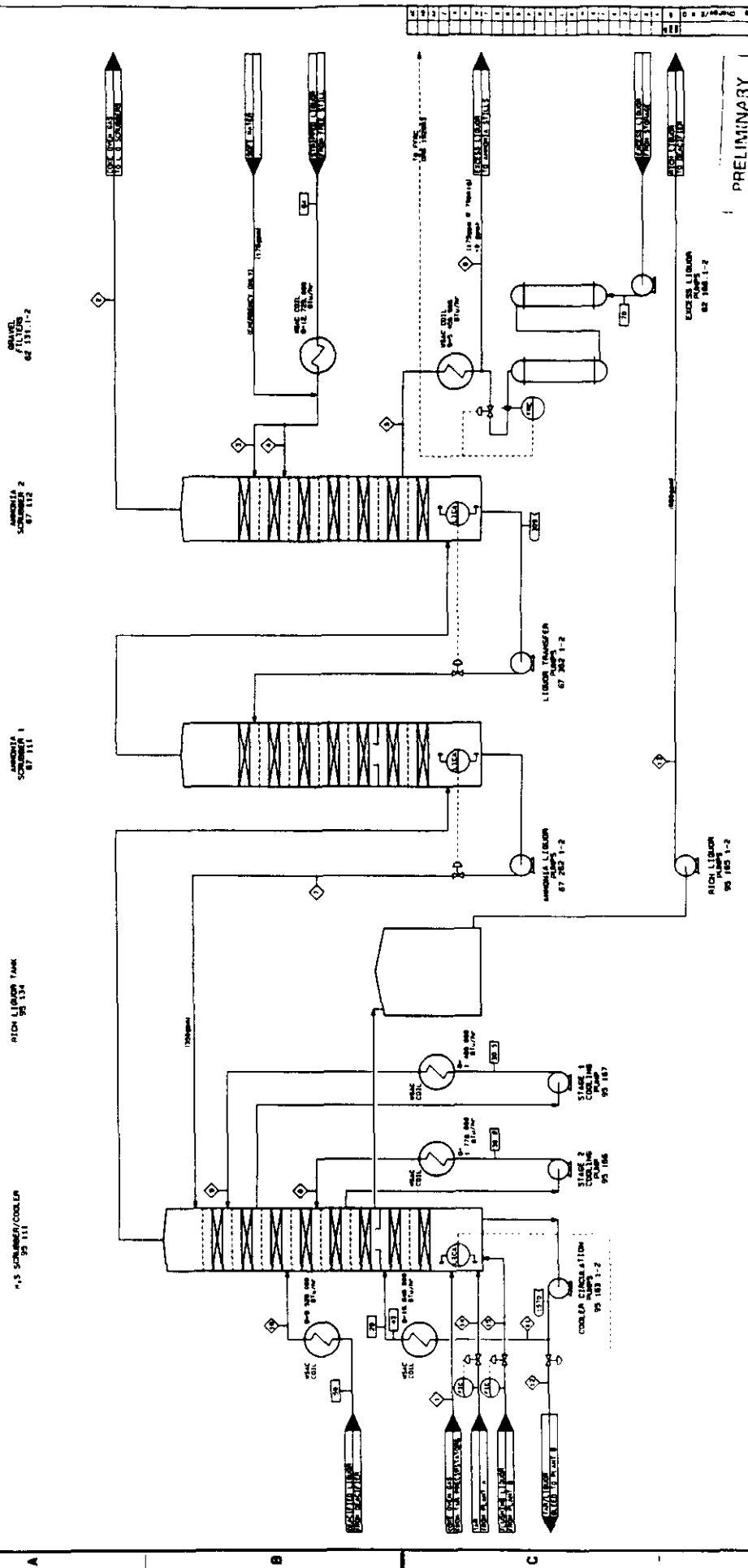
<u>Utility</u>	<u>Usage</u>
Electric Power Absorbed, KW/hr	52
LP Steam Produced (Net), lb/hr	8,800
MP Steam, lb/hr	880
City water, GPH	1,320
Natural gas, MMBTU/hr	
Normal	0
Start-up	1.7

## 3.9 PROCESS FLOW DIAGRAMS

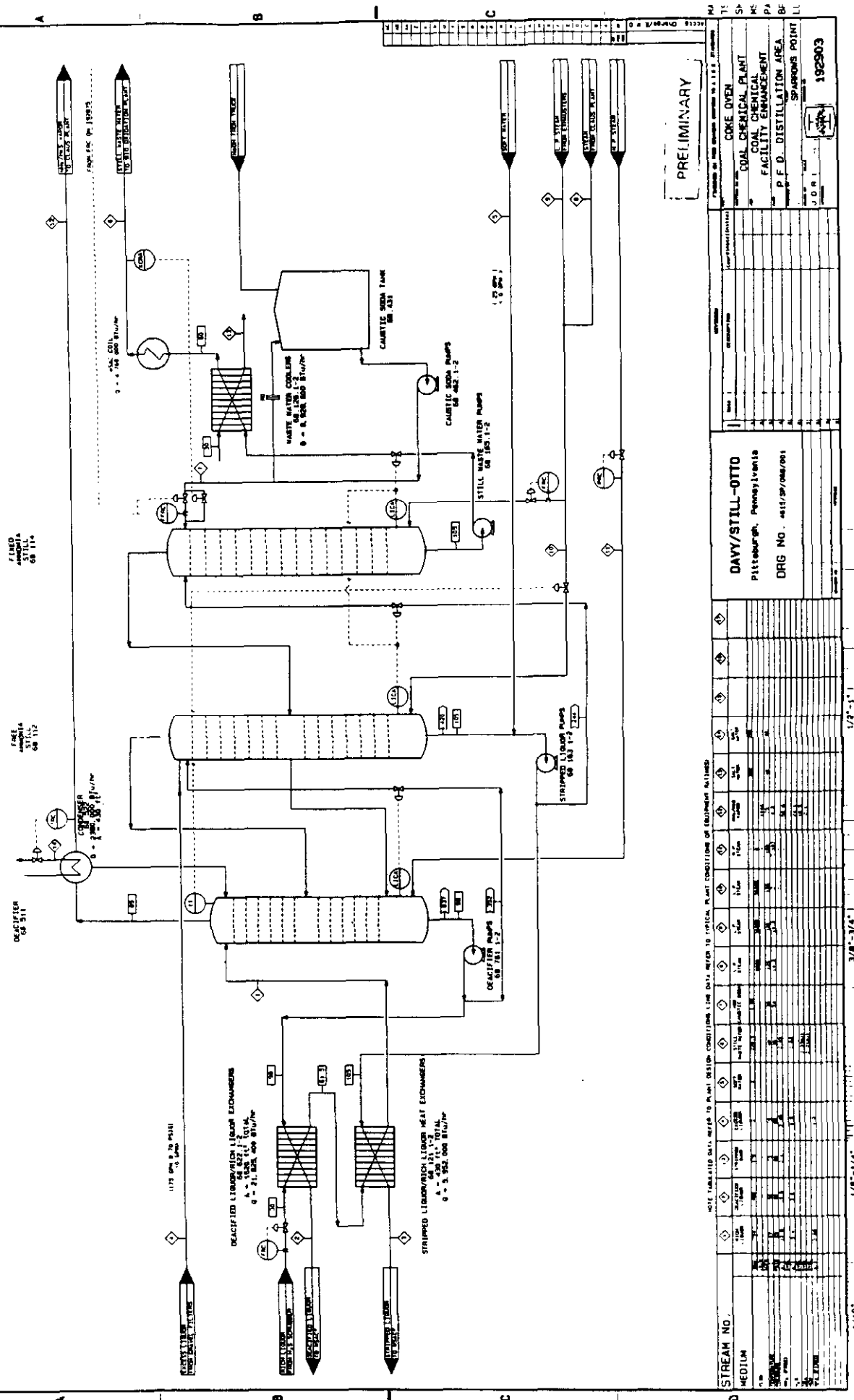
Figures 3.9.1-1, 3.9.1-2, and 3.9.1-3 represent the process flow diagrams for the Innovative Coke Oven Gas Cleaning System. The process areas represented by each diagram are as follows:

<u>Figure</u>	<u>Description</u>
Figure 3.9.1-1	Scrubber Area - PFD 192875
Figure 3.9.1-2	Distillation Area - PFD 192903
Figure 3.9.1-3	Ammonia Destruction and Claus Plant - PFD 193476

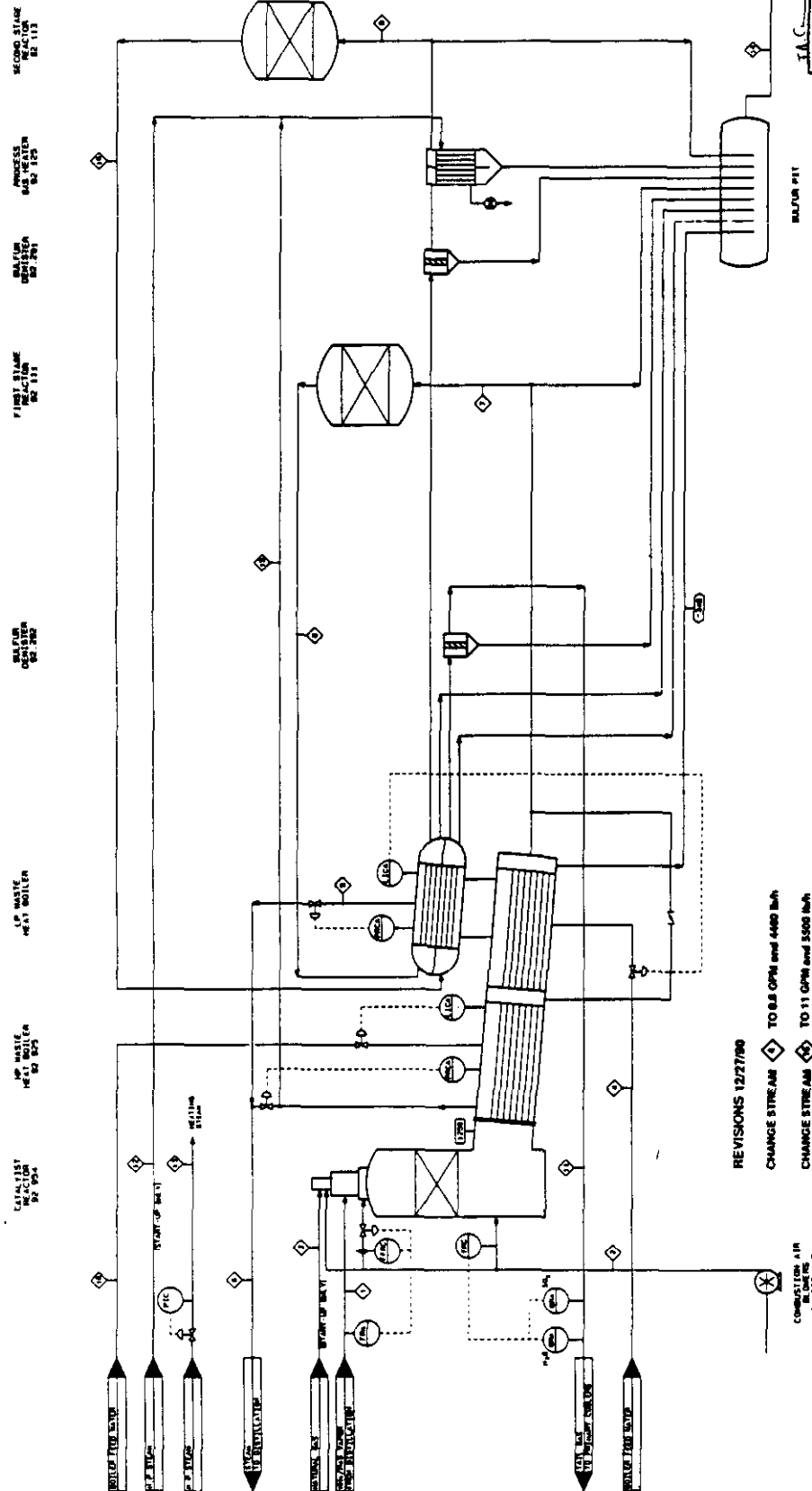
**FIGURE 3.9.1-1**

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**FIGURE 3.9.1-2**



**FIGURE 3.9.1-3**



**PRELIMINARY**

## RESULTS

REVISIONS 12/27/90

**CHANGE STREAM**  **TO 0.8 GPM and 4400 L/h**

**CHANGE STREAM**  **TO 11 GPM and 3500 lph**

THESE TABLES ARE TO BE USED IN CONNECTION WITH THE DATA REPORT TO THE NATIONAL ACADEMY OF SCIENCES

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#### 4.0 EHSS CONSIDERATIONS

EHSS impacts associated with the coke oven gas cleaning system are expected to be minimal. The consequences of both construction and operation of the project on various environmental conditions are discussed as follows:

##### Atmospheric Impacts

Although removal of existing equipment and construction and installation of new equipment generated fugitive emissions of dust at the project site, any impacts from these fugitive dusts were temporary and ceased when construction was completed. Once the system is in operation, sulfur dioxide emissions from burning coke oven gas are expected to decrease from approximately 8,900 tons per year to 2,600 tons per year. Particulate emissions are also expected to decrease since all of the coke oven gas will now undergo the same treatment. Nitrogen oxide emissions will not be significantly different since the gas cleaning system will not significantly alter the nitrogen content or heating value of the product coke oven gas. Fugitive VOC emissions will also decrease due to replacement of existing equipment with new project equipment.

No significant increase in noise emissions is expected to occur, since the project is replacing currently operating equipment. Noise resulting from the project should be indistinguishable at the plant from noise generated by other processes and equipment at Sparrows Point. There should be no discernable increase in off-site ambient noise levels.

##### Land Impacts

Construction activities occurred primarily in the area of the "B" Coal Chemicals Plant. Less than 10 acres were required for the project, and the project site is well within existing plant boundaries. No land outside plant boundaries was required for the project. All necessary utilities are currently in place.

Accepted "best management practices" were used to control stormwater run-off. Paving and curbing were installed in the process areas to collect stormwater for treatment in the existing wastewater treatment system. This satisfied the requirement to reduce the intensity and pollutant load in run-off from the project site.

Since the project eliminates the generation of low-grade ammonium sulfate, no solid wastes are produced on a routine basis. Spent catalysts will be generated approximately every five to eight years when 320 cubic feet of nickel catalyst (5 tons) and 650 cubic feet of alumina catalyst (10 tons) require replacement. The nickel catalyst is expected to be returned to the vendor for regeneration or metal recovery. Spent alumina catalyst from the old Claus plant has been found to be nonhazardous, and has been disposed of in an approved on-site landfill.

### Water Quality Impacts

Wastewater discharges from the revamped coal chemicals plant are expected to consist of process effluents, noncontact cooling water, and stormwater run-off. Process effluents will be routed to the existing biological treatment system. Loadings are expected to be lower than those from the existing plant. Non-contact cooling water requirements are expected to decrease by 24 percent after installation of the project, thereby reducing thermal discharges to the Patapsco River.

The new coke oven gas cleaning facility will use an almost completely closed-cycle process that will generate little liquid waste. The primary waste stream will be the stripped wastewater from the ammonia stills, containing low concentrations of ammonia and hydrogen sulfide, which will flow to the plant's existing biological treatment facility. This flow is projected to be 220.5 GPM, compared to the old estimated flow of 219 GPM. The existing light oil recovery unit will continue to discharge a 35 GPM liquid waste stream to the biological treatment facility. The old discharge from the cyanide stripper will be eliminated by the new process.

The wet surface air cooler and the waste heat boiler will generate 110 GPM of blowdown water containing dissolved solids. Occasional discharges from the overflow seal pots on each unit of the system will generate small intermittent flows to the treatment facility. All of these flows are small compared to the capacity of the treatment facility which is designed to handle wastewater flows up to 1422 GPM. The pollutant loading to the existing biological treatment system will also be reduced by the project. Since the old plant met discharge limits, the reduced pollutant loadings should ensure continued compliance.

### Ecological Impacts

Construction of new components and removal of old equipment temporarily disturbed several areas within the 8.6-acre project site. The entire area surrounding the project site was already highly disturbed and has been altered through slag fill and construction, with no vegetative or natural communities remaining. Other higher quality habitat areas, including wetlands, are at sufficient distance (1,500 feet) from the project site that they were not impacted by the construction.

No threatened or endangered plants or animals are present on or near the project site. In addition, there are no expected negative impacts to water and air quality, with some beneficial consequences in terms of decreases in pollutant loading to the biological treatment system and sulfur dioxide emissions to the atmosphere. Therefore, no negative impacts to biota in the area are expected.

### Socioeconomic Impacts

Construction labor was approximately 385,000 man-hours spread over a 22-month period. Operational manpower requirements did not change significantly from normal levels. Due to the availability of an ample work force in the metropolitan Baltimore area, the project had negligible effect on population and local employment. Therefore, there were no impacts to local housing, support facilities or transportation.

### Energy and Materials Impact

All necessary utilities are already in place at the plant. River water, electricity, and natural gas requirements will decrease, while steam requirements will remain unchanged. The requirements for potable water, which is supplied by the City of Baltimore, and industrial water, which is treated sewage effluent, will increase slightly. Both types of water are available in sufficient quantities to serve the project. Sodium hydroxide and other chemicals are also required and are readily available.

## 5.0 PROJECT SCHEDULE

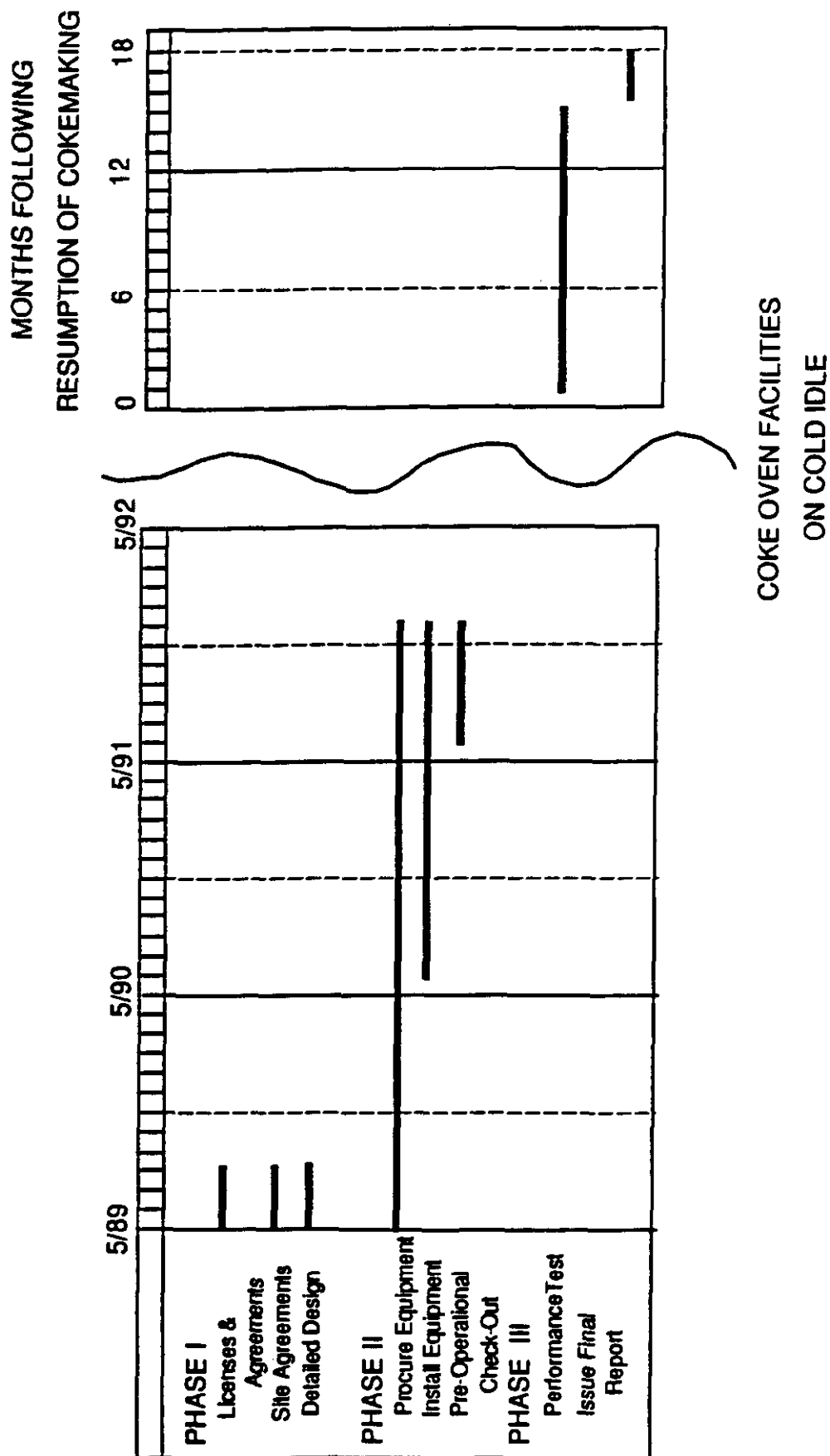
Environmental permits and preliminary design were essentially complete prior to the execution of the Cooperative Agreement. Project construction began in February 1990 and took twenty-two months to complete. Operations were scheduled to continue for fourteen months after startup. Bethlehem Steel Corporation was then to continue operation beyond the period of DOE cost sharing. The critical project tasks are identified and scheduled as shown in Figure 5.0-1.

Consistent with Bethlehem Steel Corporation's announcement in September of 1991, the production of coke at the Sparrows Point Plant stopped in December of 1991. The coke oven batteries were allowed to "go cold" in January of 1992. The coke oven gas cleaning demonstration facility was prepared for an extended shut-down of at least two years to maintain the facility for a successful start-up if and when coke-making operations are resumed. Bethlehem Steel Corporation is continuing to study a variety of options to resume coke-making at the Sparrows Point Plant. However, it appears unlikely that the coke oven batteries will be operational in the immediate future. Such a decision would be subject to business conditions, availability/demands of funds for other capital projects, and other factors.

# Figure 5.0-1 PROJECT SCHEDULE

CLEAN COAL PROJECT

SPARROWS POINT, MD.



## 6.0 COST

### 6.1 PROJECT COSTS

The financing of the Innovative Coke Oven Gas Cleaning System Project is provided by Bethlehem Steel Corporation along with financial assistance from DOE under Cooperative Agreement No. DE-FC22-90PC89658.

The budgeted cost including 14 months operation of the project per the Cooperative Agreement with DOE is \$45,239,781. A breakdown of this total is shown in Table 6.1-1.

### 6.2 CAPITAL COSTS

A breakdown of the actual mechanical equipment costs by functional area is shown in Table 6.2-1.

### 6.3 STARTUP COSTS

A breakdown of the estimated startup costs is shown in Table 6.3-1.

### 6.4 OPERATING AND MAINTENANCE COSTS

A breakdown of the estimated operating and maintenance costs is shown in Table 6.4-1.

TABLE 6.1-I

Budgeted  
Project Cost Breakdown

Innovative Coke Oven Gas Cleaning System

<u>Phase</u>	Budget per Cooperative <u>Agreement</u>
I	\$735,000*
II	\$33,684,836
III	<u>\$10,819,945</u>
Total	\$45,239,781

\* Includes Pre-Award of \$682,000

TABLE 6.2-1

Plant Cost Summary

Actual Capital Costs  
Innovative Coke Oven Gas Cleaning System

Mechanical Equipment:	\$7,170,000
Structural Steel:	\$1,220,000
Piping (includes Scrubber gas mains):	\$4,060,000
Instruments:	\$1,950,000
Electrical:	\$320,000
Construction:	
Civil:	\$ 4,290,000
Mechanical & General:	\$17,310,000
Electrical/Instruments:	\$ 2,940,000
Bethlehem Construction Support Forces:	\$ 2,201,000
Indirects:	
D/SO; BSC; ETA and Comstock	<u>\$ 6,540,000</u>
TOTAL	\$48,001,000

TABLE 6.2-1

Mechanical Equipment CostsSecondary Cooling/H<sub>2</sub>S Removal

H <sub>2</sub> S Scrubber/Cooler	730,000
Rich Liquor Tank	125,000
Pumps	70,000
Miscellaneous Equipment	15,000
Subtotal	940,000

## Ammonia Scrubbers

Ammonia Scrubber 1	340,000
Ammonia Scrubber 2	340,000
Pumps	30,000
Miscellaneous Equipment	30,000
Subtotal	740,000

## Wet Surface Air Cooling

Wet Surface Air Coolers (2)	1,015,000
Pumps	55,000
Subtotal	1,070,000

## Deacidification and Distillation

Deacidifier (2)	1,115,000
Free Ammonia Still (2)	490,000
Fixed Ammonia Still	240,000
Exchangers	185,000
Pumps	110,000
Standby Thermal Oxidizer	265,000
Deaerator & Water Softening Unit	55,000
Miscellaneous Equipment	110,000
Subtotal	2,570,000

TABLE 6.2-1 - Continued

## Sulfur Recovery

Claus Reactors (2)	60,000
Sulfur Condenser	185,000
HP Waste Heat Boiler	105,000
LP Waste Heat Boiler	145,000
Reaction Vessel	100,000
Process Gas Heater	90,000
Pumps	50,000
Miscellaneous Equipment	155,000
Boiler Feedwater System	50,000
Subtotal	940,000

## Miscellaneous Equipment

Gravel Filter Vessels	140,000
GF Tanks, Pumps, Air Blower	65,000
Tar Pumps	15,000
Catalyst	130,000
Miscellaneous	260,000
Spares	300,000
Subtotal	910,000

Total Mechanical Equipment Costs	7,170,000
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TABLE 6.3-1

Start-up Costs

Initial Supply of Consumables	\$50,000
Operations Start-up Project Team	\$400,000
Facility Start-up and Shakedown	\$400,000
Davy/Still-Otto Consultants	<u>\$ 83,000</u>
	\$933,000

TABLE 6.4-1

Projected  
Operating and Maintenance Costs\*

PHASE III Demonstration Plant Operation for 14-month period:

Task:

I	Training	\$83,000
II	Shakedown	939,000
III	Plant Operation	9,008,000
IV	Technical Support	0
V	Plant Reassessment	0
VI	BSC Program Management	50,000
VII	BSC Project Engineering	0
VIII	BSC Support Engineering	740,000
IX	BSC Operations	0
	Subtotal Phase III	\$10,820,000

\*as submitted to DOE June 6, 1989

## 7.0 REFERENCES

- [1] Comprehensive Report to Congress Clean Coal Technology Program, Innovative Coke Oven Gas Cleaning System for Retrofit Applications, A Project Proposed by Bethlehem Steel Corporation, U.S. Department of Energy, Assistant Secretary for Fossil Energy, Office of Clean Coal Technology, Washington, DC 20585, DOE/FE-0137, August 1989.
- [2] "Latest Desulfurization and Ammonia Removal/Destruction Technology for Bethlehem Steel, Sparrows Point" R. H. Schrader and D. B. Evans, Bethlehem Steel, and S. P. Haywood and M. Platts, Davy/Still Otto, Annual Association of Iron and Steel Engineers Conference, September 1989.
- [3] Environmental Assessment, Innovative Clean Coal Technology Program Coke Oven Gas Cleaning Demonstration Project, Bethlehem Steel Corporation, Sparrows Point Plant, Baltimore County, Maryland, U.S. Department of Energy, Assistant Secretary for Fossil Energy, DOE/EA-0404, December 1989.
- [4] Final Environmental Information Volume for the Coke Oven Gas Cleaning Project at the Bethlehem Steel Corporation Sparrows Point Plant, DOE/PC/89659-T1, Environmental Resources Management, Inc., April 24, 1990.
- [5] Bethlehem Steel Corporation, Contract Specification No. 327-18389 for Modification/Integration "A & B" Coal Chemical Plants.
- [6] "Cleaning Up Coke-Oven Fuel Gas", 33 Metals Producing, Page 26, February, 1992.
- [7] "Bethlehem Steel's Innovative Cleanup System Nears Completion:", Clean Coal Today, Issue No. 3, Pages 4-5, Summer, 1991.
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